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Heat and Mass Transfer Data for Condensing Pure and Binary Vapor Mixturesof Paraffinic and Aromatic Hydrocarbons.

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HEAT AND MASS TRANSFER DATA FOR CONDENSING PURE AND BINARY VAPOR MIXTURES OF PARAFFINIC AND AROMATIC HYDROCARBONS

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ABSTRACT

Methods and data for applying Nusselt's equation to the condensation of binary vapors have not been satisfactorily established. No data are available on heat-and mass-transfer coefficients for condensing binary hydrocarbon vapors which form a single-phase condensate; the vapor-liquid interface temperature needed in calculating these coefficients is still subject to definition, although limited experimental data have indicated that the bubble point of the condensate may be taken as the interface temperature. The purpose of this study is to obtain experimental data that could be used to answer these and other questions quantitatively, so that more satisfactory methods can be established for the design of condensers.

Condensation data were collected by varying the temperature drop across the condensate film covering the range from 10 to 35°F; a constant composition of condensate was maintained during each series of tests. Systems studied were: benzene-n-heptane, n-heptane-toluene, and n-hexane-toluene; the mixtures tested covered a wide concentration range of each.

Heat-transfer coefficients were calculated using as the temperature difference, that between the condensing surface: and the vapor temperature, the dew point and the boiling point of the condensate, and the interfacial temperature. The coefficients based on the interfacial temperature and the bubble point of condensate were usually intermediate between the coefficients for the pure components when the two temperatures were close, as would be expected. However, those based on the vapor

temperature and the dew point were not.

A study was made also to determine the manner in which the heat-transfer coefficients varied with composition, for various constant temperature differences. Coefficients for binary systems can be predicted with satisfactory accuracy at relatively high temperature drops, based on the bubble point of the condensate, by assuming a linear variation of coefficients with composition (expressed as mole percent) between the values of the pure components at the same temperature drop. The departure from linearity, greatest at low temperature drops, may be related to the fractionation efficiency of the condensation, whether the condensation is total, equilibrium, or between these two types.

Interpretating the data in terms of mass transfer indicated that appreciable resistance to heat and mass transfer existed in both the vapor and liquid phases. For most mixtures the liquid-phase resistance to mass transfer was more than 50 percent of the total resistance, which is contrary to the past assumption that it was absent. The smaller gas-phase resistance indicated that most of the mass was transferred bodily rather than by diffusion.

When the liquid-phase resistance to mass transfer was not controlling the interface composition was almost the same as the condensate composition and the calculated interfacial temperature was closer to the bubble point of the condensate than to either the dew point or the vapor temperature. However, when the resistance was appreciable, composition of the condensate and the interface differed markedly as did the two

temperatures.

Enrichment of the residual vapors was 30 to 80 percent of that predicted by the equilibrium diagram for the condensate composition. The higher the ratio of the individual gas-phase resistance to the over-all gas-phase resistance, the higher the degree of enrichment. Also the enrichment was directly related to the number of over-all gas-phase transfer units—which varied from 0.4 to 1.5. However, only at low enrichment does a given percentage increase in the number of transfer units effect approximately the same increase in enrichment.

The over-all height of a gas-phase transfer unit, H_{og}, varied with composition from 0.4 to 1.6 Ft. For the three binary systems, H_{og} showed a minimum value and became large as the binary mixtures became concentrated in either components.

CHAPTER I

INTRODUCTION

In most industrial surface condensers wherein the condensing fluid is an organic compound, the resistance of the condensed liquid-film may be greater than any or all other resistances involved. Consequently, designers of condensers found it very important to be able to predict what the resistance of this film of condensate would be. The problem was attacked from both the theoretical and experimental approaches; and results were most satisfactory when the condensing fluid was a single pure compound.

Essentially all of the mathematical correlations and equations which are used today for predicting condensing film coefficients of heat transfer on the outside of horizontal tubes are based on work of Nusselt (71) who first formulated the mathematics involved in condensing pure vapors. Although some of the assumptions made by Nusselt (see page 7) in deriving his equation were challenged by later investigators (9, 10, 17, 20, 45, 52, 72, 81), the equations have been verified qualitatively quite frequently by work on pure compounds. Quantitative results of different investigatiors generally agree with the corresponding theoretical values to ± 30 per cent depending upon the design of the equipment and the operating methods employed. Nusselt's final equation for condensing pure vapor on the outside of a horizontal tube is presented in the two following forms:

$$h_m = 0.725 \left(\frac{k_f^3 \rho_f^2 g \lambda}{\mu_f D_o (T_v - T_w)} \right)^{1/4}$$

or $$\frac{h_m}{\phi} = 1.51 \left(\frac{4\Gamma}{\mu_f} \right)^{-1/3}$$

In these equations all of the physical properties except latent heat of condensation, λ , are determined at a film temperature proposed by Drew (62) which is presented in Equation 2.

In the case of condensation of binary vapors forming miscible liquids, two resistances to heat transfer occur; that of the liquid-film and that of a laminar layer of vapor between the vapor-liquid interface and the main vapor stream. It has been assumed that the theoretical equation for liquid condensate film of pure components, would hold equally true for the liquid-film of a binary system. Attempts have been made in the past by several investigators to apply Nusselt's equation to the case of condensing mixed vapors. While several investigators (2,3,4,52) have studied the condensation of binary vapors forming a two-phase condensate. However, a search of the literature shows that the experimental data are very meager for the systems forming a single-phase condensate (75,113). A theoretical approach was proposed by Colburn and Drew (21) for including the resistance of the gas-film but no data were published which could test the theoretical equation. This approach was concerned primarily with the evaluation of temperature drop in the gas-film in terms of both heat and mass transfer coefficients. Their final equation was somewhat cumbersome and should be solved by trial and error.

This problem can be simplified and the rate of heat transfer predicted by use of the liquid coefficient alone if the temperature at the vapor-liquid interface can be predicted. Three simple possibilities are reported (21,75,113) as the temperature to be used as vapor-liquid temperature: (1) the vapor temperature measured directly, (2) the dew point of the vapor, and (3) the bubble point of the condensate. The last two temperatures can be read from the temperature-composition diagram for the system. Based on available experimental data, reportedly (75) the bubble point of condensate appears to be the best choice.

Condensation of mixed vapors has been investigated at the Louisiana State University. The binary systems studied prior to 1955-1956 when this project was undertaken included methanol-isopropanol (105), methanol with acetone and benzene (104), and benzene with ethanol and n-buthanol (59). In these investigations it was assumed that the liquid-phase resistance to mass transfer was absent and the project was concerned mainly to verify the correct temperature to use as the interface temperature when condensing binary vapors, one or both of the two components of which were alcohols. In the present work, a study of the binary mixtures of aromatic and paraffinic hydrocarbons was made. The primary reason for the selection of the hydrocarbon mixtures was that no previous work has been done on the hydrocarbon systems. Also these binary mixtures are completely ideal systems, versus azeotropic or non-ideal systems studied to date.

In selecting the binary mixtures, the following points were considered:

(1) Availability of vapor-liquid equilibrium data for the system.

(2) Non-azeotropic systems.

(3) Components with widely different theoretical (calculated) heat transfer coefficients.

(4) Components with large differences in boiling points and refractive indices.

(5) Components which would not have any adverse effects on the equipment in use.

On the above basis the systems selected were, (1) benzene-n-heptane, (2) n-heptane-toluene, and (3) n-hexane-toluene.

The objectives of the project were:

(1) To check the validity of the Nusselt's assumptions in condensing pure and mixed vapors of these components.

(2) To determine the limitations on the use of the boiling point of the condensate as the interfacial temperature and to determine what temperature to use when this was not the correct one.

(3) To determine improved methods for predicting the condensing heat transfer coefficients of the binary vapors from those of the constituents pure components

(4) To obtain mass transfer data; in particular: (a) To determine the individual and over-all gas-phase mass transfer coefficients and to check the validity of the assumption that liquid-phase resistance to mass transfer is absent in condensation of binary vapors. Should the assumption prove to be incorrect, to establish the

criterion of the gas and the liquid-film resistances,

(b) To determine the height and the number of individual and the over-all gas-phase transfer units and their relationship with the composition of the binary systems,

(c) To determine the degree of enrichment of the residual vapors and to correlate this enrichment with the resistances to heat and mass transfer and the number of transfer units.

CHAPTER II

THEORETICAL CONSIDERATIONS

Condensation of Pure Vapors

As soon as a saturated vapor touches a surface which is colder than itself condensation will commence, and the liberated heat will be transferred to the cooling surface. At the same time a film of condensate is formed on the cooling surface and further latent heat transferred must pass through this film. The thickness of this film influences the rate of heat transmission, since the heat can be transferred only by conduction through the film if it is laminar, and the thermal conductivity of the liquids encountered in practice is small. The film will be thicker the more slowly the condensate drains from the surface, and hence will tend to increase in thickness with increasing deviation of the surface from the vertical position. On the other hand, the thickness of the film will increase toward the bottom by drainage from the above. The velocity of drainage must also be a function of the viscosity of the condensate. The lower the viscosity, the more rapidly the condensate will drain, and the thinner the film will be. Since for all liquids encountered in practice the viscosity decreases with increase in temperature, the surface conductance for condensing vapor must, under otherwise constant conditions, increase with increase in temperature. These considerations refer to vapors without any appreciable velocity. If the vapor has a definite velocity it will influence the flow of the fluid in the film.

The relations just outlined were discussed and developed in mathematical form by Nusselt (71) in 1916. He developed his equations on the basis of the following assumptions:

1. The heat delivered by the vapor is latent heat only.
2. The condensate film is drained from the surface by laminar flow only, and the heat is transferred through the film only by conduction.
3. The film thickness at any point is a function of the mean velocity of flow and of the amount of condensate passing at that point.
4. The velocity of the condensate layer is a function of the relation between frictional shearing force and the weight of the film. Drainage is due to gravity only.
5. The amount of condensate is proportional to the heat transferred, which in turn is related to the thickness of the film and temperature difference between condensing vapor and the surface.
6. The film of condensate is so thin that the temperature gradient through it is linear.
7. The physical properties of the condensate may be taken at the mean film temperature.
8. The surface is relatively smooth and clean.
9. The curvature of the film may be neglected.
10. The temperature of the solid surface is constant.

By combining the laws of laminar flow of a fluid and of heat of conduction through it, Nusselt arrived at the following equation for condensing pure vapors on horizontal tubes

$$h_m = 0.725 \left(\frac{k_f^3 \rho_f^2 g \lambda}{\mu_f D_o (T_V - T_W)} \right)^{1/4} \qquad 1$$

Detailed derivation of this equation is given in several references (46,51,52,62,65).

The Nusselt derivation assumes that the physical properties of the condensate are constant throughout the layer, whereas the presence of a temperature gradient means that there is some variation in these properties, even though small in many cases. Based upon assumption of a linear temperature gradient through the film and a linear variation of fluidity, $1/\mu_f$, with temperature, Drew (62) derived the following equation for the mean condensate film temperature and found that the original Nusselt's equations hold true if the physical properties are evaluated at such film temperature,

$$T_F = T_V - 0.75 (T_V - T_W) \qquad 2$$

Equation 1 can be written in the following form to include mass rate of flow of condensate, W'.

$$h_m = 0.952 \left(\frac{k_f^3 \rho_f^2 g L}{W' \mu_f} \right)^{1/3} \qquad 3$$

Equation 3 can also be written in a dimensionless form to include Reynolds number, $4\Gamma/\mu_f$ where Γ is condensing load = W'/L, W' being the mass rate of flow, L the length of the horizontal tube.

For a horizontal tube,

$$\text{Re. No.} = \frac{4r_h G}{\mu_f} = \frac{4}{\mu_f}\,\frac{S}{L}\,\frac{W'/2}{S} = \frac{2W'}{\mu_f L} = \frac{2\Gamma}{\mu_f} \qquad 4$$

Substituting $L/W' = 2/\text{Re.No.}\ \mu_f$ from the above equation in Equation 3, and rearranging

$$h_m = 0.952 \left(\frac{k_f^3 \rho_f^2 g}{\mu_f} \; \frac{2}{\text{Re.No.}\ \mu_f} \right)^{1/3}$$

$$h_m \left(\frac{k_f^3 \rho_f^2 g}{\mu_f} \right)^{-1/3} = 0.952\ (2)^{1/3}\ (\text{Re.No.})^{1/3} \qquad 5$$

$$\frac{h_m}{\phi} = 1.2\ (\text{Re.No.})^{-1/3} \qquad 6$$

where

$$\phi = \left(\frac{k_f^2 \rho_f^2 g}{\mu_f^2} \right)^{1/3} \qquad 7$$

Since a Reynolds number of 2100 can theoretically be reached on each side before turbulence begins, it follows that the critical values of $4\Gamma/\mu_f$ for the entire horizontal tube is 4200. Therefore, Equation 6 can take the form of

$$\frac{h_m}{\phi} = 1.51\ (4\Gamma/\mu_f)^{-1/3} \qquad 8$$

Detailed discussion of numerous experimental data for pure materials is available in literature. McAdams (62) gives a summary of the data of

many observers condensing pure saturated vapors outside single horizontal tubes.

Some of the observers have made modifications to Nusselt analysis. Peck (72) believed that the two basic assumptions made by Nusselt—that the temperature drop across the condensing film is constant around the tube and that the condensate film is not subject to acceleration forces—were not valid. Through a series of equations, a semiempirical equation was developed which gave values of the heat transfer coefficient which were not dependent of the above two assumptions. Bromley, et al. (9) points out the magnitude of the integral

$$\frac{4}{3}\int_{o}^{\alpha} \sin^{1/3}\alpha \, d\alpha \qquad 9$$

used in the derivation of Equation 1, re-evaluated by means of gamma functions, to be 3.4495, which results in a constant 0.728 instead of 0.725 in Equation 1. Bromley, et al. (9) investigated both theoretically and experimentally the error introduced by the assumption of a constant tube temperature around a horizontal condenser tube. Therefore, a term was introduced to account for the effect of temperature variation around a horizontal condenser tube. However, both Peck (72) and Bromley (9) concluded that conduction of heat around a tube has a negligible effect on the over-all heat transfer coefficient. Equation 1, becomes more complicated when the temperature profile through the film is not assumed to be linear and the effect of cross-flow—flow perpendicular to the

condensing surface—within the film is considered. For these considerations, the following correction factor to Equation 1, is recommended by Rohsenow (81) as an improvement over an earlier contribution by Bromley (10),

$$\left(1 + 0.68 \quad \frac{C_{pf}\ \Delta T}{\lambda}\right)^{1/4} \qquad 10$$

Here C_{pf} is the specific heat of the condensate at the mean film temperature, T_F, and λ is the latent heat of condensation of the condensate.

This correction factor, Equation 10, is recommended (81) as a satisfactory approximation to the more complex equations proposed by Bromley (10) for the range of

$$o < \frac{C_{pf}\ \Delta T}{\lambda} < 1.0 \qquad 11$$

It is believed that the effect of this refinement becomes more important at high values of liquid subcooling or more precisely high values of $C_{pf}\ \Delta T/\lambda$. The magnitude of $C_{pf}\ \Delta T/\lambda$ for great many cases is quite small (0 to 0.2). For such cases Rohsenow (81) proposed the latent heat term λ used in Equation 1 be modified to λ', where

$$\lambda' = \lambda + \frac{3}{8} C_{pf}\ \Delta T \qquad 12$$

However, for cases where $C_{pf}\ \Delta T/\lambda$ is smaller than 0.2, such as

hydrocarbons used in the present work, the magnitude of the λ' does not differ greatly from λ and its effect upon theoretically calculated h_m is less than two per cent.

The effect of vapor velocity outside a tube and at right angles to the drainage was not calculated by Nusselt. However, Jakob, Erk and Eck (45) found the condensing coefficient of steam inside a vertical tube to vary directly with the velocity of vapor. Schmidt (91) condensed CO_2 inside a vertical tube at a pressure up to the critical point. The experimental values were about twice Nusselt's values and showed a marked dependence on vapor velocity. Monrad and Badger (65) noted that of much greater importance than the effect of vapor velocity was the effect of turbulence in the fluid film of the condensate. Kirkbride (53) reported that the main reason that experimental data on condensing vapors on vertical tubes were higher than Nusselt's values, was the effect of the turbulence in the condensate film. Colburn (20), reviewed the results of Kirkbride and developed a correlation equation based on a simplified description of the condensate layer which was considered as being composed of a laminar sublayer and a turbulent outer region. The laminar layer was considered to have the controlling resistance to heat transfer. This type of analysis was extended by Carpenter and Colburn (17) to include the effect of shear stress. Rohsenow and co-workers (82) present analyses showing the effect on rate of condensation of vapor shear stress at the liquid-vapor interface. Both laminar and turbulent films were considered and were combined to give analytical results for the case of laminar flow on the upper portion of a vertical plate and the turbulent

flow on the lower portion.

Condensation of Binary Mixed Vapors

To apply the Nusselt film coefficient of heat transfer to the condensation of mixed vapor, it is necessary to determine the temperature drop across the condensing film. This requires the evaluation of the temperature at the vapor-liquid interface. Colburn and Drew (21) have studied this problem by relating mass and heat transfer. Their work is summarized in the following paragraphs:

Figure 1 shows the local conditions at a point on the cooling surface of a condenser operating under steady state conditions. The vapor mixture, with a composition y_{AV} of the more volatile component A, flows at a temperature T_V parallel to the surface of the wall, and is separated from that face by a stream of condensate in which the mole fraction of component is x_c.

The local heat flux per unit area Q absorbed by the cooling medium is the sum of three quantities: q_c, the sensible heat of cooling the condensate; $q_{\lambda I}$, the latent heat evolved by condensation at the interface I - I; and q_s, the sensible heat received from the vapor. All of these q's are considered for unit area of the vapor side of the wall. Usually q_c and q_s are small compared to $q_{\lambda I}$. The picture is the same as that for condensing a pure vapor, and, if the interfacial temperature T_I can be determined, the heat transfer problem for the condensate layer is reduced essentially to the pure vapor case, and is thus solved.

The Lewis-Whitman film theory of mass transfer (56) infers that between the interface I - I and the main vapor stream there exist a layer

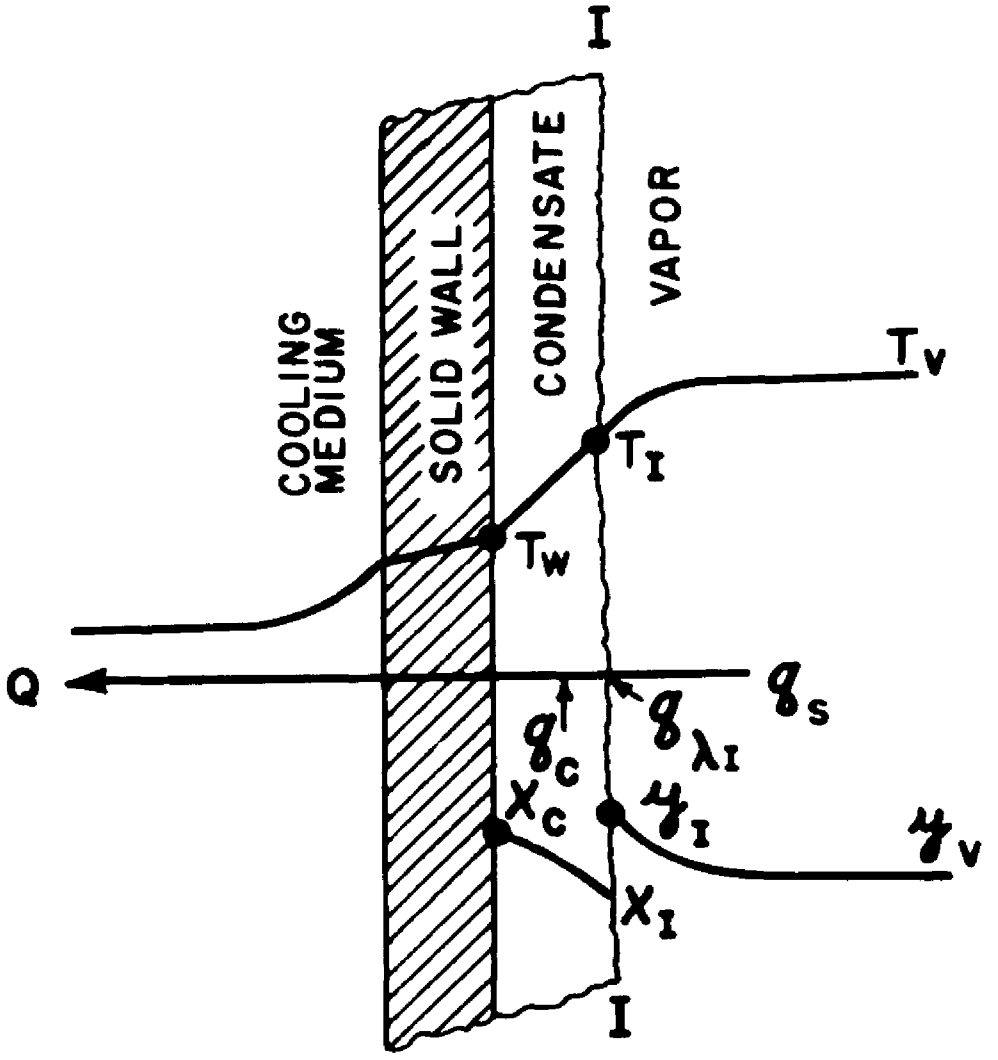


FIGURE I.

TEMPERATURE AND CONCENTRATION GRADIENTS FOR A CONDENSING BINARY VAPOR

of vapor that is in laminar flow. At points within this layer the molal rate of transfer of the lower boiling component toward the interface in the cases of equimolal counter diffusion would be given by

$$N_A = - D_V \frac{dc_A}{dz} = - \frac{D_V}{RT} \frac{dp_A}{dz} \qquad 13$$

c_A = concentration of component A, Lb.-moles/Ft.^{3}
D_V^A = molecular diffusivity in gas, Ft.^{2}/Hr.
N^A = Rate of diffusion of component A, Lb.-moles/Hr.Ft.^{2}
p^A = partial pressure of more volatile component, A, Atm.
R^A = gas constant, Ft.^{3} atm/lb-mole, °R.
T = absolute temperature, °R.
z = distance in direction of diffusion, Ft.

Equation 13 expresses the mass transfer due to molecular diffusion alone. However, another factor, eddy diffusivity, is involved which must be considered. Equation 13 indicates that molecular diffusion in the gas is proportional to the partial pressure gradient of the diffusing gas. Since eddy diffusion in a gas stream is also proportional to the partial pressure gradient, the mixed process of eddy and molecular diffusion involved in a mass transfer may be expected also to be proportional to the partial pressure gradient. Since the relative importance of eddy diffusion and molecular diffusion is not known, the unknown diffusivity and the average length of the diffusion path are combined together in the form of a coefficient k_g. Then, the molal rate of transfer of the more volatile components, A, by eddy and molecular diffusion toward the interface is defined by the following equation

$$N_A = k_g (p_{AV} - p_{AI}) \qquad 14$$

where p_{AV} is the partial pressure of the more volatile component in the main vapor stream and p_{AI} is the partial pressure of the more volatile component at the interface.

For the more general case of condensation of binary vapors with no inert gas present, the movement of mixed vapor toward the cold surface carries the components more rapidly than if transfer were solely by diffusion. Let N_A and N_B represent the molal rate of condensation of components A and B from the binary vapor mixture containing y_{AV} mole fraction of the more volatile component A at a total pressure P. Let y_{AI} be the mole fraction of component A at the interface I - I (at T_I) and let z be the distance from the bulk vapor in the direction of diffusion. Then, the mass transfer to the interface between vapor and liquid which is brought about by diffusion through the laminar film and the bodily transport due to condensation may be represented by

$$N_A = k_g Z_g \frac{dp_A}{dz} + (N_A + N_B)\, y_A \qquad 15$$

where k_g is "film coefficient" of mass transfer and Z_g is the thickness of the laminar vapor layer.

In Equation 15, the first term on the right represents the rate of diffusion of component A with respect to (A + B) and the second the rate of bodily transport of A due to condensation of A and B. The former is an approximation only, since thermal diffusion is neglected; k_g for the component A depends on the partial pressure and diffusion coefficient of component B. Equation 15 may be integrated from $y_A = y_{AV}$ at $z = 0$

(main body vapor mixtures) to $y_A = y_{AI}$ at $z = Z_g$ (condensate surface), the result is

$$(N_A + N_B) = k_g P \ln \frac{\frac{N_A}{N_A + N_B} - y_{AI}}{\frac{N_A}{N_A + N_B} - y_{AV}} \qquad 16$$

At the vapor-liquid interface, N_A and N_B are at a temperature T_I. Neglecting any heats of mixing and assuming that the enthalpies of pure components in both the vapor and liquid phases are additive, the latent heat of mixture is expressed as:

$$q_{\lambda I} = N_A (h_{gA} - h_{lA}) + N_B (h_{gB} - h_{lB}) \qquad 17$$

where $q_{\lambda I}$ is heat of condensation and h_g and h_l are enthalpies of saturated vapor and saturated liquid, respectively, both being taken at T_I. Assuming the molal latent heats of both components to be equal, Equation 17 can be expressed as:

$$q_{\lambda I} = (N_A + N_B) \lambda_I \qquad 18$$

where λ_I is the mean molal latent heat of the components at the temperature T_I.

The heat flux q_s, which arrives at the interface as sensible heat, arises from two sources: first, the cooling of main stream of vapor and, second, the cooling from the main stream temperature T_V to the interfacial

temperature, T_I, of the $(N_A + N_B)$ lb.-moles/Hr.Ft.^{2} of the vapor which are condensing on the area under consideration. Assuming no condensation takes place in the film, the amount of sensible heat arriving at the interface is

$$q_s = h_g \left[Z_g \left(\frac{dT}{dz} \right) \right] + (N_A M_A C_{pA} + N_B M_B C_{pB})(T_V - T_I) \qquad 19$$

In this equation M and C_p are respectively, the molecular weight and specific heat of the components. The first term in this equation represents the heat flux arising from local temperature gradient and the second term measures the heat accompanying the bulk flow of condensing vapor stream. Since only temperature varies with Z_g in Equation 19, it integrates into

$$q_s = h_g (T_V - T_I) \frac{C_o}{1 - e^{-C_o}} \qquad 20$$

where

$$C_o = \frac{N_A M_A C_{pA} + N_A M_B C_{pB}}{h_g} \qquad 21$$

Equation 21 can be reduced to

$$C_o = \frac{G C_{pf}}{h_g}, \qquad 22$$

where G is $(N_A M_A + N_B M_B)$ and C_{pf} is the weighted average specific heat of vapor in the gas film.

Substitution of Equation 16, in Equation 18, for $(N_A + N_B)$ gives

$$q_{\lambda I} = \lambda_I \, k_g \, P \ln \frac{\frac{N_A}{N_A + N_B} - y_{AI}}{\frac{N_A}{N_A + N_B} - y_{AV}} \qquad 23$$

The sum of Equations 20 and 23 is the total of heat flux delivered to the condensate surface (21).

$$q_s + q_{\lambda I} = \frac{C_o \, h_g}{1 - e^{-C_o}} (T_V - T_I) + \lambda_I \, k_g \, P \ln \frac{\frac{N_A}{N_A + N_B} - y_{AI}}{\frac{N_A}{N_A + N_B} - y_{AV}} \qquad 24$$

The values of h_g and k_g may be evaluated from the following relations (94,106),

$$J_H = \frac{h_g}{C_p G} \left(\frac{C_p \mu}{k} \right)^{2/3} \qquad 25$$

$$J_D = \frac{k_g P}{G_M} \left(\frac{\mu}{\rho D_V} \right)^{2/3} \qquad 26$$

where G and G_M are mass rate of flow and molal rate of flow, respectively.

Johnstone and Pigford (47) report data on the rectification of several binary mixtures in a wetted wall column, under conditions such that the gas-film resistance was 90 per cent or more of the total

resistance to mass transfer. Their data is presented by Sherwood and Pigford (94) as:

$$J_D = 0.033\ Re^{-0.23} \qquad 27$$

Treybal (106) has tabulated J_H and J_D relations for various situations in heat and mass transfer operations. He also shows the J_H and J_D factors graphically as a function of Reynolds number or its modified form, depending upon flow conditions and whether the operation is heat or mass transfer. For flow of gases transverse to cylinders, Treybal recommends Equations 25 and 26, except that the exponents 2/3 in both equations are changed to 0.56, which shows a better correlation than the 2/3 power. Using experimental data of other investigators, Sherwood and Pigford (94) also found that for flow of gases transverse to cylinders, the exponent 0.56 would correlate the experimental data better than when the exponent 2/3 is used.

Equation 16 can be written in exponential form as follows:

$$\frac{x_A - y_{AV}}{x_A - y_{AI}} = e^{-\left(\frac{N_A + N_B}{k_g P}\right)} \qquad 28$$

where $x_A = \frac{N_A}{N_A + N_B}$ is the ratio of the rate of condensation of the more volatile component to the net rate of condensation—that is, the mole fraction of the more volatile component in the condensate. Substituting k_g from Equation 26 in Equation 28,

$$\frac{x_A - y_{AV}}{x_A - y_{AI}} = e^{-\left(\frac{(N_A + N_B)(P)(\mu/\rho D_v)^{2/3}}{J_D G_M P} \right)} = E \qquad 29$$

Rearranging Equation 29, gives

$$x_A = \frac{y_{AV} - E\, y_{AI}}{1 - E} \qquad 30$$

The values of y_{AV} and y_{AI} are between zero and unity and if condensation is taking place, $(N_A + N_B)$ is positive and E has a value between zero and unity.

When the condensate rate is very small and the value of E approaches unity, Equation 29 shows that y_{AI} would approach y_{AV}; thus, the condensate present is in static equilibrium with the main vapor stream. At the other extreme, at large condensation rates when the value of E is zero or approaches zero, Equation 30 shows that x_A would equal y_{AV}, or in short, bulk flow controls the mass transfer.

Thus, the temperature at the interface is related to the value of E. When E is equal to or approaches zero and the composition of the condensate approaches that of the main vapor stream, the temperature at the interface would approach or equal that of main vapor stream provided there is no superheat. On the other hand, when E is unity, then the condensate is in static equilibrium with vapor and the temperature of the interface is that at which the binary mixture of that composition would exist at the given pressure, that is, at its normal boiling point. Thus, the limits between which the interfacial temperature could vary

would be the bubble point of a solution and the dew point of the vapor of the same composition. Making use of mass transfer data, the temperature of the interface could be established between the above limits. With the temperature at the interface known, the temperature drop across the film could be evaluated for use in Nusselt's equation.

Using Colburn and Drew (21) nomenclature for $N_A/(N_A + N_B) = z$, for component A, the more volatile component, Equation 16 is modified to:

$$w\,z = N_A = k_g\,z \ln \frac{z - y_{AI}}{z - y_{AV}} \qquad 31$$

Assuming V and L represent the molar mass rates of vapor and liquid, respectively, Kent and Pigford (50) derived the following equation

$$w\,z = \frac{-d\,(y_{AV}V)}{dA} = \frac{-d\,(x_A\,L)}{dA} \qquad 32$$

$$\text{or} \quad w\,(z - y_{AV}) = -V\,(dy_{AV}/dA) \qquad 33$$

Here y_{AV} and x_A represents the mole fraction of component A, the more volatile components, in the vapor phase and liquid phases, respectively, and dA represents a differential element of heat transfer area. Kent and Pigford considered an idealized system in which it was assumed that there was no liquid-phase resistance to mass transfer and that the interfacial resistance was absent. Under these assumptions, the authors combined Equations 31 and 33 to get:

$$V\, d\, y_{AV} = (y_{AI} - y_{AV}) \frac{dV}{1 - e^{w/k_g}} \qquad 34$$

Equation 34 can be written as

$$\int_{y_{A1}}^{y_{A2}} \frac{dy_A}{y_{AI} - y_A} = \int_{V_1}^{V_2} \frac{d(\ln V)}{1 - e^{w/k_g}} \qquad 35$$

where for simplicity the subscript V is eliminated from y_{AV}.

The integral on the left of Equation 35 is frequently called "the number of transfer units because its numerical value is a measure of the amount of composition change accomplished in a binary system, regardless of its particular phase-equilibrium characteristics" (50). If y_{AI} is considered constant, the integral can be solved for as shown here:

$$\int_{y_{A1}}^{y_{A2}} \frac{dy_A}{y_{AI} - y_A} = -\int_{y_{A1}}^{y_{A2}} \frac{-dy_A}{y_{AI} - y_A} = \Big[-\ln (y_{AI} - y_A) \Big]_{y_{A1}}^{y_{A2}}$$

$$= -\ln (y_{AI} - y_{A2}) + \ln (y_{AI} - y_{A1})$$

$$= \ln \frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} = N_g \qquad 36$$

Numerical evaluation of the integral on the right of Equation 35

"shows the influence of partial condensation on the enrichment. In order to evaluate it precisely the functional dependence of k_g on the local value of V should be allowed for" (50). Although exact numerical values can be found, Kent and Pigford report that it is sufficiently precise to assume that k_g is proportional to V and to use the logarithmic-mean value of k_g in the denominator of the integral. The authors propose the following method of obtaining a solution to the integral

$$\left[\frac{w}{k_g} \right]_{Avg.} \doteq \frac{(G_{M1} - G_{M2})(\pi D^2/4\pi Dh)}{(k_g)_{l.m.}} \qquad 37$$

For the particular apparatus (condensation inside a vertical pipe) used in their investigation, the interfacial area per unit volume of apparatus, a, is given as

$$a = \pi D h / \frac{\pi}{4} D^2 h \qquad 38$$

$$a h = 4 \pi D h / \pi D^2 \qquad 39$$

where D and h are the diameter and the height of the column used in their investigation. Substituting Equation 39 in Equation 37,

$$\left[\frac{w}{k_g} \right]_{Avg.} = \frac{G_{M1} - G_{M2}}{(k_g)_{l.m.}} ah = \frac{G_{M1} - G_{M2}}{\frac{k_{g1} - k_{g2}}{\ln \frac{k_{g1}}{k_{g2}}} ah} \qquad 40$$

$$\text{Now, } k_{g1} = \frac{k_{g1}}{G_{M1}} \; G_{M1} \tag{41}$$

$$\text{Assuming } \frac{k_{g1}}{G_{M1}} = \frac{k_{g2}}{G_{M2}} \quad , \tag{42}$$

$$k_{g2} = \frac{k_{g1}}{G_{M1}} \; G_{M2} \tag{43}$$

Substituting Equations 41 and 43 in 40,

$$\left[\frac{w}{k_g} \right]_{Avg.} = \frac{G_{M1} - G_{M2}}{\dfrac{\dfrac{k_{g1}}{G_{M1}} G_{M1} - \dfrac{k_{g1}}{G_{M1}} G_{M2}}{\ln \dfrac{k_{g1}}{G_{M1}} G_{M1} \dfrac{G_{M1}}{k_{g1} G_{M2}}}} \; ah$$

$$= \frac{G_{M1} - G_{M2}}{\left(\dfrac{k_{g1}}{G_{M1}} \right) \dfrac{G_{M1} - G_{M2}}{\ln \dfrac{G_{M1}}{G_{M2}}}} \; ah$$

$$= \frac{\left(G_{M1} - G_{M2} \right) \left[\ln \dfrac{G_{M1}}{G_{M2}} \right] G_{M1}}{k_{g1} \left(G_{M1} - G_{M2} \right) ah} \tag{44}$$

$$\left[\frac{w}{k_g}\right]_{Avg.} = \frac{\left[\ln \frac{G_{M1}}{G_{M2}}\right] G_{M1}}{k_{g1}\, ah} = \ln (G_{M1}/G_{M2}) \frac{H_{g1}}{h} \qquad 45$$

$$\text{where } H_{g1} = \frac{G_{M1}}{k_{g1}\, a} = \frac{V_1}{k_{g1}\, aS} \qquad 46$$

Substituting V_1/V_2 for G_{M1}/G_{M2}, Equation 45 is modified to

$$\left[\frac{w}{k_g}\right]_{Avg.} = \ln (V_1/V_2) \frac{H_{g1}}{h} \qquad 47$$

With these approximations, Equation 35 may be solved (50) as:

$$N_g = \int_{y_{A1}}^{y_{A2}} \frac{d\, y_A}{y_{AI} - y_A} = \frac{d(\ln V)}{1 - e^{w/k_g}}$$

$$\doteq \frac{\ln (V_1/V_2)}{1 - (V_1/V_2)^{-H_{g1}/h}} \qquad 48$$

Combining Equations 36 and 48,

$$\ln \frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} = \frac{\ln (V_1/V_2)}{1 - (V_1/V_2)^{-H_{g1}/h}} \qquad 49$$

If the term H_{g1}, the composition and the molal mass flow rates of vapors entering and leaving the system are known, the vapor composition at the interface, y_{AI}, can be calculated from Equation 49. Once y_{AI} is known, the interfacial temperature T_I can be read from the temperature-composition diagram. Knowing T_I, the condensing wall temperature, T_W, the total heat flux, q, and the condensing surface, A, the Newton's equation

$$q = h_I A \Delta T_I \qquad 50$$

can be solved to determine h_I, the film coefficient of heat transfer.

When the term $(V_1/V_2)^{-H_g/h}$ in the denominator of Equation 49, is sufficiently small so that the denominator is essentially unity, Equation 49 simplifies as

$$\ln \frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} = \ln (V_1/V_2) \qquad 51$$

or

$$\frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} = V_1/V_2 \qquad 52$$

Applying Equation 14 to condensation of a binary vapor mixture, the rate of transfer of a single component (here the more volatile component) is given by Pressburg (77) as:

$$N_A A = (V_1 y_{A1} - V_2 y_{A2}) - (V_1 - V_2) y_{A1}$$

$$= V_2 (y_{A1} - y_{A2}) = k_g A (y_A - y_{AI})_{l.m.} \qquad 53$$

In the above equation, the net quantity of the first parentheses in the left-side is the gross disappearance of the more volatile component from the vapor phase. This is due both to condensation of vapor and to diffusion of the more volatile component relative to the bulk of flow. The term $(V_1 - V_2)\ y_{A1}$ indicates that if there were no diffusion of the components, then the disappearance of the more volatile components would be bulk flow which is the change in volume of the vapor times its concentration. Thus the net quantity of the left-side of Equation 53, which simplified to $V_2\ (y_{A1} - y_{A2})$, should be the net transfer of the more volatile component by diffusion alone. The equation meets the test of the two extreme limits. First for case of total condensation, where no fractionation is taking place and thus the composition of the vapor entering and leaving the apparatus does not change, the quantity of N_A will be zero. Second for rectification, where there will not be any change in volume of vapor entering and leaving the apparatus, the rate of transfer may be expressed by $N_A\ A = V\ (\triangle\ y_A)$ which is accepted value.

Knowing the required compositions and vapor flow rates, one can solve Equation 53 for experimental values of k_g.

The number of individual gas-phase transfer units may be determined from Equation 36.

The height of an individual gas-phase transfer unit may be determined from the following equation (94).

$$h = H_g\ N_g \qquad 54$$

or

$$H_g = h/N_g \qquad 55$$

This also can be determined from,

$$H_g = \frac{G_M}{k_g a P} \qquad 56$$

Of course, determination of H_g by the two Equations 55 and 56 should give the same results and these equations are not an independent check on each other. This can be easily understood from the following equation

$$H_g = \frac{h}{N_g} = \frac{G_M}{k_g a P} \qquad 57$$

where both N_g (Equation 36) and k_g (Equation 53) are based on the same driving force. In fact, Equation 57 shows that H_g, calculated in both ways, must check.

The foregoing equations lmay also be written in terms of over-all gas-phase driving force. For example, Equation 53 may be expressed, in terms of over-all gas-phase driving force and over-all gas-phase mass transfer coefficient, as

$$V_2 (y_{A1} - y_{A2}) = K_g A (y_A - y_A^*)_{l.m.} \qquad 58$$

which can be solved for K_g.

Similarly,

$$h = H_{og} N_{og} \tag{59}$$

or

$$H_{og} = h/N_{og} \tag{60}$$

The number of over-all gas-phase transfer units may be determined from the following integral

$$N_{og} = \int_{y_{A1}}^{y_{A2}} \frac{d\, y_A}{y_A^* - y_A} = \ln \frac{y_A^* - y_{A1}}{y_A^* - y_{A2}} \tag{61}$$

The individual and over-all gas-phase resistances may be correlated (94) as

$$\frac{1}{K_g} = \frac{1}{k_g} + \frac{m}{k_l} \tag{62}$$

where m is the slope of the equilibrium line which may be expressed as

$$m = \frac{d\, y_A^*}{d\, x_A} \tag{63}$$

The height of individual and over-all gas-phase transfer units may be related (94) to each other by

$$H_{og} = H_g + m \frac{G_M}{L_M} H_l \tag{64}$$

where H_l is the height of a liquid-phase transfer unit and m is the slope of the equilibrium line which is expressed by Equation 63.

Similar equations may be written in terms of individual or overall liquid-phase driving force. However, since such equations will not be used in the present study, their derivation and explanation will be omitted here.

CHAPTER III

PREVIOUS EXPERIMENTAL WORK

A search of the literature shows that very little work has been done in determining the heat transfer coefficient for condensing mixed vapors forming a single-phase condensate. The only experimental data published is that of Wallace and Davison (103). Their investigation was undertaken to determine film coefficients of heat transfer for the liquids condensing from mixed vapors in the case of a simple binary system. The system selected was ethanol-water and condensation was effected on the outside of a horizontal tube.

The condensing surface used was a 30 inch long brass tube with an outside diameter of 0.84 inches placed inside a 3 inch diameter steel jacket. Seventy to ninety per cent of the vapor entering the test section was condensed. Uncondensed vapors from the test section were passed into a glass auxiliary condenser, where they were totally condensed. Mercury-in-glass thermometers were employed to measure the temperature of entering and leaving cooling water and the vapor. The surface temperature was measured by 6 copper-constantan thermocouples imbedded on the test surface and arithmetic average of these readings were taken as the average surface temperature. The binary mixture placed in a steel drum was heated and vaporized with a steam coil. Condensate was collected and weighed and a heat balance was made on the basis of the quantity of the cooling water and the temperature raise.

The initial experiments were carried out with the composition of

the condensate kept constant and the cooling water rate varied. Later work was performed allowing the condensate composition to vary with constant cooling water rate.

The determined heat transfer coefficients were plotted against the cooling water flow rate for four different compositions of condensate from mixture of alcohol-water. The film coefficient decreased with an increasing velocity of cooling water. This effect became more pronounced as the composition of mixed vapors deviated from pure alcohol. It was, therefore, unsatisfactory to apply Wilson reciprocal plot, which assumes that the coefficient for condensing film to be independent of water rate, to system of mixed vapors.

Heat transfer coefficients were also plotted against the concentration of the condensing mixture for two constant flow rates. Based on these water rates, the following equation was derived which relates the condensing heat transfer coefficient, h, to the mean mole fraction of ethanol, M, in the condensate.

$$h = (M - 0.008)\ e^{(11.45 - 8.35\ M)} + 350 \qquad 65$$

The relationship was found to be satisfactory for mole fraction of ethanol above 0.10.

It was assumed that temperature drop across the liquid film would be the difference between the temperature of vapor and the tube surface. The coefficient calculated in this manner was called the "apparent" coefficient. "Actual h" values were calculated on the assumption that the true temperature at the liquid-vapor interface was the boiling point of the condensate and not the actual vapor temperature. In order

to illustrate the differences between apparent and actual coefficients, the two values were plotted on the same graph against mole fraction ethanol in the condensate. There is little deviation for high mole fraction of ethanol, but when the water content in the condensate increased above 50 per cent, the differences between the two curves became increasingly greater.

In summary, all the data obtained represented high condensation rates and were in agreement with the Colburn-Drew (21) prediction that the condensate composition would approximate that of the entering vapor.

Other data on film coefficients for condensing binary vapors have been obtained by Trachtenberg (105), Todd (104), and Malone (59). Trachtenberg investigated the methanol-isopropanol system. Todd experimented on acetone-methanol and methanol-benzene systems. Malone studied ethanol-benzene and n-butyl alcohol-benzene systems. All the data were obtained with the apparatus used in the present work, which except for slight modification is that originally built by Todd (103). Their investigation was carried out essentially in the same manner as used in the present work which is fully explained in Chapter V. For each concentration the calculated heat transfer coefficients were plotted against temperature drop across the film. Using values from these curves, plots were made of film coefficients at constant temperature drop against the composition of the condensate expressed as mole percent. The curves obtained for the coefficients with temperature drop across the film based on the bubble point of the condensate were more linear than those using a temperature difference across the film based on either the vapor

temperature or the dew point of the condensate. However, even these curves were not completely straight line; in an attempt to smooth out these curves, Todd (104) plotted the bubble point coefficients against weight per cent composition of the more volatile component in the condensate. This was accomplished somewhat, especially in the case of the methanol-isopropanol system which approaches ideality more so than the other systems investigated.

The curves based on the dew point and the vapor temperature were also similar in appearance and such should be the case for the two temperatures were not greatly different. In all cases these latter curves exhibited a minimum point. The temperature-composition diagrams showed that these minimums occurred where the differences between the bubble point and dew point of the condensate were greatest.

From the results of these studies it was concluded that the bubble point of the condensate was the correct temperature to use with the theoretical Nusselt equation. It was also concluded that, for the systems investigated, the coefficients can be predicted without any serious error by linear interpolation of the values of the pure components at the same temperature drop across the film.

Kent and Pigford (50) primarily interested in fractionation effected in condensation, reviewed the problem of the simultaneous condensation of a binary vapor mixture and, in addition, obtained experimental evidence to show that mass transfer effects are in agreement with the Colburn-Drew theory. The authors report the results of experimental studies of a condenser used to fractionate mixtures of ethylene

dichloride and toluene. Details of the apparatus used in their investigation are given in reference (50). Very briefly, the apparatus consisted of a wetted-wall distillation column surrounded with an annular jacket. Water was circulated in the annular jacket, under controlled conditions, to produce known and nearly uniform rate of condensation. With an external reflux, supplied from a total condenser operating on the exit vapor stream from the test section, the column could be operated as an adiabatic counter current distillation unit or as a partial condenser.

The investigation showed that the observed mass transfer effects were in agreement with the Colburn-Drew film theory, which accounts for the influence of the net mass exchange on the relative mass transfer rates of the components. It was found by Kent and Pigford that observed resistances to mass transfer agreed qualitatively with values independently measured; the diffusional resistance of the liquid-phase had a pronounced effect on fractionation, which conformed to the Rayleigh theory of equilibrium differential condensation.

CHAPTER IV

APPARATUS

Briefly, the apparatus consisted of a steam heated reboiler connected to a vapor jacket surrounding a 3.908" diameter copper bar, mounted horizontally, which acted as a condensing surface. The vapor jacket was connected to an auxiliary condenser to take care of any vapor not condensed on the primary condensing surface. Cooling water was introduced into both ends of the copper bar through especially drilled holes and removed at the opposite ends. A system was provided for the collection of condensate from a known condensing surface. The rate of heat transferred was determined by using the weight of the condensate and the latent heat data. The temperature drop was obtained by measuring the temperature of the vapor and of the condensing surface. A schematic diagram and a photograph of the system are shown in Figures 2 and 3, respectively.

To determine an accurate film coefficient, it was necessary to provide a precise method for evaluating the amount of heat transferred, q, and the temperature drop across the film of condensate, Δt, for a given heat transfer area, A. Of these terms, probably the most difficult to measure accurately was the temperature drop across the film of condensate. It was very much desirable that its value be known very accurately, and it be of the same magnitude for all points on the condensing surface. For the pure liquids, the temperature drop across the film was the difference between temperature of the saturated vapor and the temperature of the condensing surface, whereas for the binary

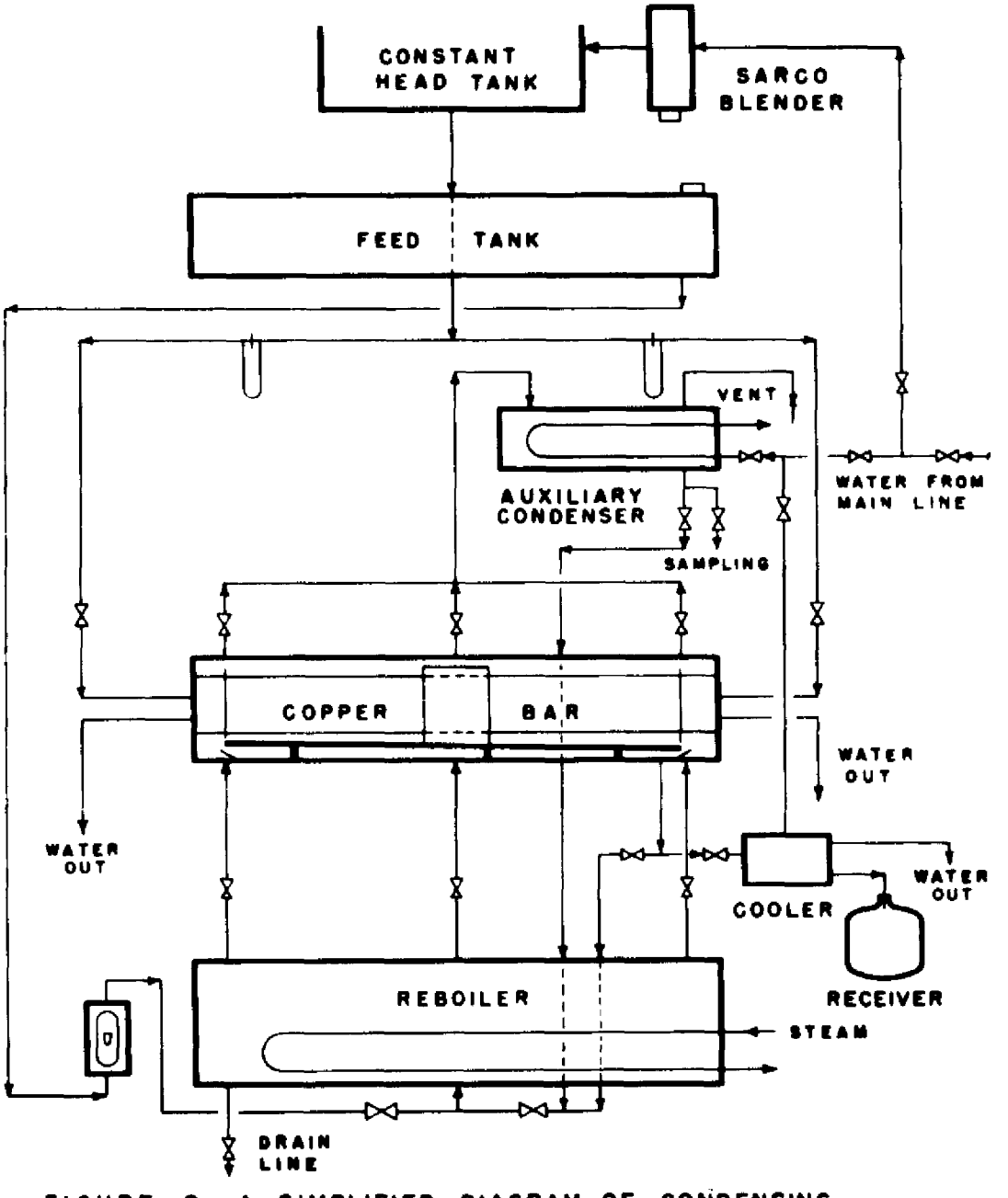


FIGURE 2. A SIMPLIFIED DIAGRAM OF CONDENSING HEAT TRANSFER APPARATUS

FIGURE 3. A VIEW OF THE APPARATUS

mixtures it was desired to determine the correct temperature difference across the condensate film.

For a complete detail of the design and structure of the unit reference is made to work of Todd (103,104). Several modifications made for the purpose of this particular study are described; however. For the present work the reboiler system was modified to include a Diaphragm Control Valve, Type 86-D, accompanied by a Type 1440 Pressure Controller from Black, Sivalls and Bryson, Inc., and a Turnbuckle type Valve Positioner from Moore Products Company. With the new control valve it was possible to control the heat input to the system to maintain a pressure of 2" to 4" of water in the vapor jacket throughout a run. The outlets from the steam coils were put under a liquid seal provided by the condenser water. Upon Malone's suggestion (59) a double bottom was furnished for the condensate trough. The space between the double bottom was filled with asbestos powder to act as an insulant to prevent any reboiling of condensate when systems involving components of widely different boiling points are studied. This is shown in Figure 4.

Originally, a mercury thermometer was placed in the vapor jacket. Upon Malone's suggestion, the thermometer was replaced by four copper-constantan thermocouples of 30 B and S gauge. Four 3/8" holes, two on the top and two on the sides, were drilled on the vapor jacket and tapped for 1/8" pipe thread. The thermocouple wells were made of silver soldered 1/4" copper tubing and were extended 1/2" in the vapor jacket. Figure 5 shows the location of the thermocouple holes on the jacket. To measure the temperature of the condensate, a copper constantan

FIGURE 4

SCHEMATIC DIAGRAM SHOWING LOCATION OF THE DOUBLE BOTTOM COLLECTING TROUGH IN THE PRIMARY CONDENSER

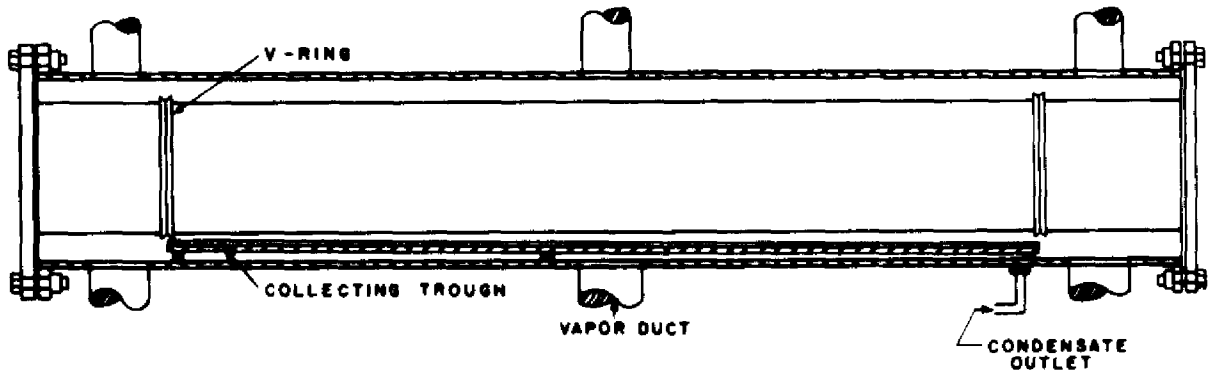
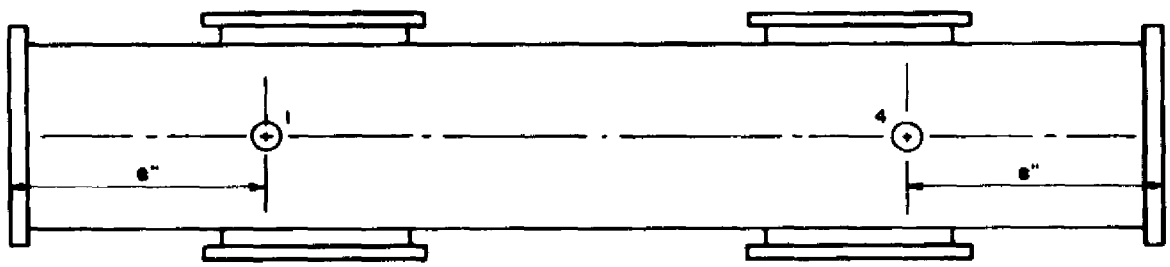
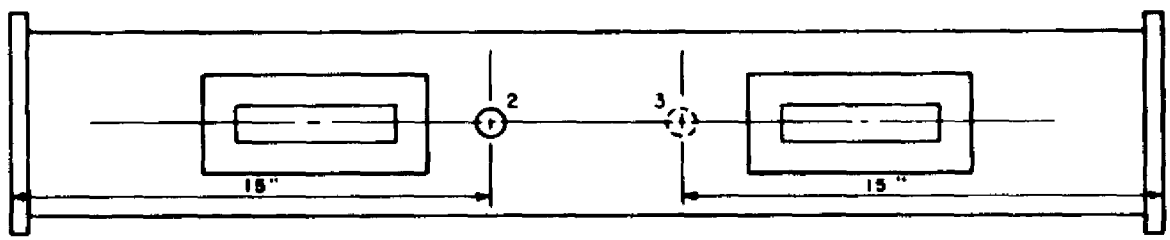


FIGURE 5

LOCATION OF 4 THERMOCOUPLE HOLES ON THE JACKET AND SCHEMATIC DIAGRAM OF A THERMOCOUPLE WELL



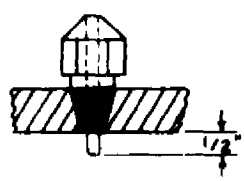
TOP VIEW



SIDE VIEW

NOTE

NOT TO SCALE
JACKET DIAMETER 6".
JACKET LENGTH 36".
HOLES DIAMETER 1/2"



was placed in a 1/4" copper tubing-well that was welded into the condensate line.

The temperatures indicated by four thermocouples installed in the vapor jacket and the one put in the condensate line were checked against a series of known temperatures. For this purpose, distilled water and lubricating oil were used as heating medium. In each test the cold junctions of the thermocouples were kept in ice-water bath at 32°F. and the hot junctions were immersed with a thermometer in the heating medium. The medium was heated to a fixed known temperature and the e.m.f. generated by each thermocouple was recorded. The medium was heated further to a higher temperature and then its temperature dropped until it reached the fixed known temperature and the generated e.m.f. were recorded again. The temperature readings for the five thermocouples are summarized in Table I. The highest temperature difference noted between the thermometer readings and the thermocouple readings is 1.8°F for heating lubricating oil at 180°F. The discrepancy between the thermometer and the thermocouple readings can be attributed to either or both of these causes: that the thermometer used was neither a standard nor a corrected thermometer; or the heating medium was not and could not be kept at a constant temperature. The thermometer indicated a change as high as 0.5°F in the bath temperature. With these limitations in mind, it can be said that the readings shown by the thermocouples are accurate enough for this work and the thermocouples can be used safely without any calibration. Temperature was measured by reading the electromotive force generated by the thermocouples using a Brown

TABLE I

VARIATION OF TEMPERATURES INDICATED BY THE JACKET THERMOCOUPLES

(Cold Junction: Distilled Water at 32°F.)

Thermocouple No.	Temperature Reading of the Thermocouples, °F.									
	1		2		3		4		5	
Temperature of the Medium °F.	Heating	Cooling	Heating	Cooling	Heating	Cooling	Heating	Cooling	Heating	Cooling
	Heating Medium: Distilled Water									
100	99.2	101.0	100.0	100.0	99.8	100.4	100.6	100.5	100.0	99.8
120	118.9	120.6	120.3	119.3	119.5	119.5	120.2	119.2	120.3	120.5
140	139.9	140.0	139.4	139.8	140.4	140.5	140.2	140.3	141.0	140.1
160	159.2	160.0	159.4	159.6	159.3	159.8	160.0	160.0	160.2	160.4
180	180.4	179.5	181.3	179.7	180.5	179.6	180.0	180.3	180.6	179.7
200	199.1	199.6	199.4	199.6	200.0	199.7	199.2	199.5	200.1	200.7
	Heating Medium: Lube Oil									
80	80.3	80.0	79.6	80.1	79.7	79.6	79.2	79.9	80.1	80.2
100	98.8	99.3	99.2	99.8	99.2	99.1	99.6	99.7	100.0	100.4
120	120.6	118.6	120.7	119.6	119.6	118.4	119.6	118.8	119.8	119.2
140	139.3	139.7	139.2	139.4	139.6	140.2	139.7	139.9	140.0	141.2
160	160.1	158.9	159.4	159.9	159.2	160.2	160.6	161.1	159.8	160.7
180	178.8	179.6	178.2	179.5	179.2	180.0	179.7	179.9	179.6	180.1
200	198.9	199.7	199.2	200.2	199.7	199.9	199.6	200.0	200.3	200.2

Electronik High Precision Multirange Indicating Potentiometer. The reference junctions of the thermocouples were maintained at 32°F. in an ice-water bath contained in a Dewar flask.

An auxiliary condenser was assembled to the vapor jacket to condense the excess vapor passed through the jacket without being condensed on the bar. A feed tank was connected to reboiler so that the condensate could be returned as necessary to permit operation over an extended period. The entrance of the condensate feed line was located at extreme left side, at the bottom of the reboiler. Later, it was felt this arrangement could produce concentration gradient from one end of the reboiler to the other and to eliminate such a possibility, the location of the line was changed to enter near the center of the reboiler.

For the analysis of the hydrocarbon mixtues, a Bausch and Lomb Precision Refractometer, model 33-45-01, was used. Water from a constant temperature bath was circulated through the refractometer to maintain the instrument at constant temperature of 25°C.

To construct a calibration curve of refractive index versus mole percent composition for each binary system, ten mixtures of known compositions were made and the refractive indices of the mixtures were measured and the calibration curve was plotted. The data are summarized and shown in the Appendix B, Section XII.

CHAPTER V

EXPERIMENTAL WORK

The previous experiments performed on the binary mixtures at the Louisiana State University were primarily on alcohol-benzene systems. For the present study it was decided to resume the study on binary systems of paraffinic and aromatic hydrocarbons, mixtures of which show non-ideal volatility characteristics. The study was also to include ideal systems of binary mixtures of paraffinic hydrocarbons. However, for the time that was available, only three binary systems of paraffinic-aromatic mixtures were studied. The binary systems chosen were n-heptane-benzene, n-heptane-toluene, and n-hexane-toluene. Primary reasons for these choices were the facts that, (1) vapor-liquid equilibrium data were available for these systems, (2) the systems were non-azeotropic type and the equilibrium diagrams showed a relatively large temperature difference between vapor and liquid lines, (3) the components of each system had widely different theoretical (calculated) heat transfer coefficients, (4) moderately large differences between the refractive indices of the pure components provided a means of analyzing the mixtures, and (5) these compounds would not have adverse effects on the equipment in use and should cause minimum permanent deposits on the condensing surface.

Reagents

Since the operation of the unit required a rather large volume of liquid and because it was felt that these would be more representative of industrial practice, available technical or commercial grade reagents

were used without additional purification. However, purity in all cases was high, as is indicated by a comparison of refractive indices with the literature values for pure hydrocarbons as is shown below.

Hydrocarbon	Refractive Index $N_D^{25°C}$	
	Literature	Experimental
Benzene	1.49792	1.49744
Toluene	1.49414	1.49218
n-Hexane	1.37226	1.37225
n-Heptane	1.38511	1.38558

The benzene and toluene were reagent grade, thiophene-free, with a purity of 99 mole per cent and 95 mole per cent, respectively. Both were purchased from the Matheson Chemical Company. n-Hexane and n-heptane were purchased from the Phillips Petroleum Company. n-Hexane was of commercial grade and n-heptane was of technical grade with a 99.5 mole per cent purity. Phillips Petroleum Company also furnished a research grade, 98.4 volume percent pure, para-xylene. Although the specification data supplied by the organization showed a negative Doctor Test, the copper bar was completely tarnished upon the first run on the unit. Consequently this hydrocarbon was eliminated.

Procedure

At the beginning, the unit was completely dismantled, modified wherever it was needed, and cleaned. The cleaning included the reboiler, jacket, condensing surface, auxiliary condenser, water tank, feed tank, and all the steam, water, and connecting lines. Except for the condensing

surface, the rust and debris were removed either by iron brushing or air blowing.

The terminals of the thermocouples on the condensing bar were relocated and arranged in the terminal box.

The condensing bar was polished with crocus cloth and soft rags.

Once the original cleaning was over, the unit was reassembled, and a small volume of benzene was put in the system, and the unit was run for a short while to check for the possibility of leaks in the system. When it was noticed that there were no leaks in the unit, the heat supply was cut off, and after cooling the benzene was drained out.

The bar was removed from the jacket and polished twice during the experimental work. Once was when a run was made on para-xylene which completely tarnished the bar (see above). The other time was after five months of operation when the bar was tarnished. At other times, after each set of runs were completed on a pure hydrocarbon or a mixture, the bar was cleaned in the jacket by using crocus cloths and soft rags. The surface was then rinsed with the liquid to be used and polished with a soft cloth. Then the unit was purged with air to remove any traces of the liquid or vapor.

About eight gallons of liquid were placed in the reboiler to begin the experiment. This volume was sufficient to cover the steam coils during operation which could be checked by the sight glass on the reboiler. It was necessary to cover the coils to prevent any superheating

of the vapor.

At least one hour prior to any run, the reference junctions of thermocouples of the condensing surface, jacket, and the reboiler were placed in an ice-water bath so that equilibrium could be reached. For small temperature drops, the blender was adjusted to give higher cooling water temperatures. Once the temperature of the cooling water from constant head tank was established and remained constant, the cooling water flow rate through the condensing bar was adjusted. This adjustment, for the desired temperature difference, was made by regulating the flow until the outlet water temperatures from both ends of the condensing bar were the same. Theoretically, the flow of the coolant should be regulated so that the manometer in each branch of the cooling system to the bar would be equalized. However, it was found that this would result in 1°F. or 2°F. temperature difference between the water outlet temperatures from both ends of the bar. Therefore, it was decided to equalize the outlet water temperatures, rather than the flow rates, though at no time were the manometers differences as high as 0.5" of Mercury.

Flow of water through the condensate sub cooler and the auxiliary condenser were also adjusted, with enough water passing through the auxiliary condenser to insure complete condensation of the vapor entering.

All valves between the reboiler and the vapor jacket and all valves between the vapor jacket and the auxiliary condenser were opened. The vent on the auxiliary condenser was also opened, under a liquid seal of the hydrocarbon under experiment. This was necessary in order to prevent

a build up of pressure in the system and to allow the vapors to purge the air from the unit before a run was started as well as to keep the air from getting into the system in case there was a temporary vacuum due to lowering of the heat input. The condensate line valves were adjusted for recycling to the reboiler. Steam was then admitted to the heating coils and the liquid vaporized. During this initial warmup period an excess amount of vapor was fed to the auxiliary condenser to force out any non-condensables from the system. For benzene runs the steam flow then was adjusted manually. Some difficulty was encountered in maintaining a constant heat input due to pressure changes in the steam supply. This difficulty was partially eliminated by "loading" the steam supply line, thereby throttling the valve, rather than permitting to alternate between open and closed. However, when the manually controlled valve was replaced by the new automatic control valve, the above difficulty was completely eliminated. The steam flow was then adjusted until the condensate film on the bar appeared to be uniform and only a small amount of vapor passed to the auxiliary condenser. The amount of condensate from the auxiliary condenser was kept at a low value, approximately ten percent of that on the primary condenser, in order to assure almost total condensation in the vapor jacket and yet to maintain the system free from non-condensables and to prevent possible infiltration of air through unknown leaks. It was found that a pressure of 2" to 4" of water, indicated by the manometer attached to the vapor jacket, within the system maintained the desirable conditions. The warmup period was about one hour, after which the surface temperatures

of the bar, jacket temperatures and others were checked to be sure that the system was in a steady state condition.

After all adjustments had been made, the valves in the condensate lines from the jacket and the auxiliary condenser were adjusted to route the condensates to the collecting bottles. Minor adjustments in heat input were necessary as the condensates were cooled, collected, and returned through the rotameter to the boiler. When steady state conditions were resumed and were indicated for a period of 30 minutes, a run of 15 to 25 minutes was made. For pure and mixed hydrocarbons, the total amount of the condensate from both the primary and the auxiliary condensers for the timed period were collected, weighed, and immediately returned to the reboiler at the same rate as it was collected. For the pure hydrocarbons this maintained a constant demand on the heat input and eliminated the possibility of the vapor surges in the vapor jacket. For mixtures this was necessary in order to keep a constant composition during the period of a test. As an aid in keeping the rate of feed-back to the reboiler equal to the amount of condensate, particularly in the case of mixtures, about 3/4 of a gallon of the liquid was maintained in the feed tank as "holdup". The feed-back rate was controlled by use of the rotameter located between the feed tank and the reboiler. Through a trial and error procedure a rate was established so that the amount of the condensate collected was equal to the feed-back. In case of mixtures, during the trial and error period, samples of the condensate were analyzed by the use of refractive indices to determine the composition of the condensate. Once the variation in the composition of

the condensate samples were less than 0.5 mole per cent, the test was started.

During the course of a test on the mixtures, three samples of the condensate from the primary condenser, one sample of the condensate from the auxiliary condenser, and one sample of the reboiler liquid were taken and analyzed immediately after the tests of the mixture were over. The analyses were made with a refractometer and a previously constructed refractive index-composition curve.

The electromotive force generated by each thermocouple was determined three to five times during a run. The average inlet and outlet water temperatures, the jacket pressure, and the steam pressure were recorded. To evaluate the surface temperature, the readings of the surface thermocouples were averaged and the temperature was obtained from a plot of the electromotive force versus temperature. The jacket temperatures, and the entering vapor temperature were determined in the same manner.

In addition to four pure hydrocarbons, three binary systems including (a) benzene-n-heptane, (b) n-heptane-toluene, and (c) n-hexane-toluene were studied. Data were collected for each binary system at composition of approximately 20, 40, 60, and 80 mole per cent of the light component.

CHAPTER VI

RESULTS AND DISCUSSION

Both Todd (104) and Malone (59) encountered difficulties in temperature measurements and keeping a constant supply of heat to the apparatus, particularly in the case of binary mixtures. For the present work these difficulties were eliminated. The temperature measurements were made accurately and very rapidly with a Brown Electronik Potentiometer and the steam supply was controlled very easily with a control valve. Both of these new features are discussed in Chapter IV. The result of adding these new features was that the experimental data proved consistent and reproducible. To insure precision in measurements, throughout almost the entire experiment each test was duplicated for the same operating conditions.

HEAT TRANSFER DATA

One of the objectives of this work was to study the effect of composition on the heat transfer coefficient of condensing binary vapors forming a single liquid phase. To accomplish this, the film coefficient of heat transfer for the pure components was measured first; then, the film coefficient of heat transfer for binary systems of various composition of condensate was determined. The film coefficient for both pure and binary vapors was calculated from Newton's equation

$$q = W' \lambda = h A \Delta T \qquad 66$$

For pure materials the rate of heat transfer, q, was determined by

multiplying the condensate rate by the latent heat of condensation of the component, which was reported in the literature (84). The temperature difference was determined by knowing the vapor temperature and the condensing surface temperature, which were both measured accurately. Knowing the area of the condensing surface, the film coefficients of heat transfer for the pure components were calculated from Equation 66.

For binary systems, a series of determinations was carried out under such conditions as to maintain a constant condensate composition and almost total condensation. The amount of heat transferred was determined in the same manner as that for condensing pure materials, except that the latent heat of condensation for each system was determined from an enthalpy-concentration diagram which was prepared before. The temperature difference was determined in four ways: the temperature of entering vapor, the dew point of the condensate, the boiling point of the condensate, or the interfacial temperature minus the average wall temperature. Then film coefficients were calculated based on each of the above temperature differences.

The required physical properties of the pure and mixed hydrocarbons, Nusselt's physical property group, vapor-liquid equilibrium data, and enthalpy-concentration data for the binary mixtures are given in Appendix B. Step-by-step sample calculations for both pure and binary vapors are given in Appendix C.

I - Pure Vapors

The results for the pure components are tabulated in Table II A through D, and they are plotted in two forms: (1) heat transfer

TABLE II

FILM COEFFICIENTS OF CONDENSING PURE VAPORS

A BENZENE

Run No.	Temperature, °F.			Condensate Gms. / Min.	Film Coefficient Btu / Hr. Ft.² °F.		Physical Property Group	Reynolds No.	h / φ	
	Vapor	Avg. Wall Temp.	ΔT	W	Experimental	Theoretical	φ	$\frac{4\Gamma}{\mu_f}$	Experimental	Theoretical
				JACKET PRESSURE: 2" - 4" WATER						
2	177.3	140.5	36.8	948.4	242	217	894.7	241	0.270	0.243
3	177.1	139.0	38.1	1000.0	246	214	893.2	253	0.276	0.239
4	177.5	140.4	37.1	988.3	250	214	894.7	252	0.279	0.239
5	176.3	135.3	41.0	1066.0	244	208	889.4	265	0.274	0.234
6	177.2	141.0	36.2	948.4	246	226	895.0	242	0.275	0.253
7	177.2	152.7	24.5	741.3	284	236	906.4	199	0.313	0.260
8	176.8	150.6	26.2	763.3	273	232	904.2	203	0.303	0.257
9	176.9	156.2	20.7	669.4	303	237	909.7	182	0.334	0.260
10	176.7	164.9	11.8	419.4	334	280	918.0	118	0.364	0.305
11	176.6	155.4	21.2	663.9	295	243	908.8	180	0.324	0.267
12	177.1	143.1	34.0	902.2	249	220	897.2	232	0.278	0.246
13	178.0	150.9	27.0	803.0	279	228	905.0	214	0.308	0.252
14	177.5	152.5	25.0	754.2	284	232	906.4	202	0.313	0.256
15	177.1	149.0	28.1	804.4	268	228	902.8	212	0.297	0.253
39	177.9	155.7	22.2	682.0	288	241	909.4	185	0.317	0.265
40	178.1	155.5	22.6	696.0	289	240	909.4	189	0.318	0.264
41	177.8	160.2	17.6	580.2	309	253	913.8	161	0.339	0.277
42	178.3	161.2	17.1	587.4	322	253	914.9	164	0.352	0.276
43	178.1	164.2	13.9	510.7	345	264	917.7	144	0.376	0.288
44	178.6	165.5	13.1	477.3	342	270	919.2	136	0.372	0.294
45	178.2	147.2	31.0	888.5	269	219	901.5	233	0.298	0.243
46	178.1	147.5	30.6	815.6	250	232	901.2	214	0.277	0.257
47	177.8	136.6	41.2	1065.0	242	209	891.2	266	0.272	0.234
48	179.9	134.9	45.0	1095.0	228	207	890.2	273	0.256	0.233
				JACKET PRESSURE: 1" - 4" MERCURY						
16	179.8	151.0	28.8	989.5	323	213	905.0	265	0.356	0.233
17	179.2	150.5	28.7	896.2	293	220	905.0	239	0.324	0.243
18	180.7	144.3	36.4	990.0	255	213	904.6	263	0.282	0.236
19	181.0	150.3	30.7	1000.0	305	212	905.4	267	0.337	0.234
20	178.9	151.0	27.9	915.5	308	217	905.5	245	0.340	0.240
21	178.7	150.4	28.3	877.2	291	221	904.7	234	0.321	0.245
22	182.7	152.9	29.8	935.0	295	217	908.3	253	0.325	0.239
23	182.2	151.3	30.9	953.0	290	215	906.6	256	0.319	0.237
24	178.6	149.8	28.8	970.0	316	215	904.0	257	0.349	0.237
25	177.5	149.2	28.3	921.0	306	216	903.0	244	0.338	0.240
26	179.7	156.5	23.2	789.0	319	229	910.8	216	0.350	0.251
27	180.5	157.6	22.9	830.4	340	221	912.0	229	0.373	0.243
28	178.4	154.0	24.4	784.0	301	221	908.0	212	0.332	0.244
29	181.2	155.8	25.4	914.0	338	218	911.2	251	0.371	0.240
30	182.4	152.7	29.7	939.0	297	217	908.0	254	0.327	0.239
31	180.6	152.5	28.3	908.7	302	218	907.1	245	0.333	0.241
32	180.1	153.6	26.5	914.0	324	218	908.1	247	0.356	0.240
33	180.8	154.3	26.5	904.0	320	219	909.0	245	0.352	0.241
34	180.8	153.9	26.9	907.3	317	219	908.7	246	0.348	0.241
35	179.1	152.1	27.0	833.7	290	225	906.4	224	0.320	0.249
36	180.9	155.4	25.5	869.3	320	222	869.3	237	0.352	0.243
37	179.6	155.4	24.2	826.2	321	226	909.8	225	0.353	0.248
38	181.6	157.2	24.4	897.3	344	218	912.0	247	0.3787	0.240

TABLE II

FILM COEFFICIENTS OF CONDENSING PURE VAPORS

B n-HEPTANE

Run No.	Temperature, °F.			Condensate Gms. / Min.	Film Coefficient Btu / Hr. Ft.^{2} °F.		Physical Property Group	Reynolds No.	h / ϕ	
	Vapor	Avg. Wall Temp.	ΔT	W	Experimental	Theoretical	ϕ	$\frac{4\Gamma}{\mu_f}$	Experimental	Theoretical
1	210.2	186.1	24.1	596.5	186	187	778.7	246	0.239	0.241
2	210.2	186.1	24.1	625.0	196	185	778.7	258	0.251	0.237
3	210.6	173.0	37.6	1047.3	210	157	775.8	415	0.270	0.202
4	210.6	173.8	36.8	1022.4	210	158	776.0	406	0.270	0.204
5	210.8	161.0	49.8	1410.6	213	142	772.0	538	0.276	0.184
6	210.6	162.2	48.4	1427.0	222	142	773.3	546	0.288	0.183
7	210.9	170.5	40.4	1173.4	219	150	775.2	465	0.283	0.194
8	210.5	181.0	29.5	816.8	209	169	777.6	332	0.268	0.218
9	210.5	187.2	23.3	665.0	215	181	779.0	276	0.276	0.232
10	210.4	181.4	29.0	764.5	199	175	777.7	311	0.256	0.225
11	210.4	180.4	30.0	821.3	206	169	777.4	333	0.266	0.218
12	210.4	188.6	21.8	622.0	216	185	779.3	259	0.277	0.237
13	210.4	188.3	22.1	630.0	215	184	779.2	262	0.276	0.236
14	210.4	194.8	15.6	458.8	222	203	780.7	195	0.284	0.261
15	210.4	195.0	15.4	470.5	230	202	780.7	200	0.294	0.258
16	210.4	195.8	14.6	456.2	236	204	780.9	194	0.302	0.261
17	210.5	171.4	39.1	1120.0	216	154	775.4	441	0.279	0.198
18	210.8	165.1	45.7	1270.0	210	148	774.0	491	0.271	0.192
19	210.5	152.5	58.0	1590.0	207	139	771.1	590	0.268	0.180
20	210.5	151.7	58.8	1614.0	207	138	770.9	597	0.268	0.179
21	210.3	196.6	13.7	451.0	248	204	781.1	193	0.218	0.262
22	210.3	196.6	13.7	448.0	246	205	781.1	191	0.316	0.262
23	210.4	188.8	21.6	593.0	207	188	779.3	247	0.265	0.241
24	210.4	189.5	20.9	596.0	215	187	779.5	249	0.276	0.240
25	210.4	170.8	39.6	1071.0	204	156	775.2	421	0.263	0.201
26	210.4	166.2	44.2	1220.2	208	150	774.2	473	0.269	0.194
27	210.4	167.6	42.8	1184.9	209	151	774.6	461	0.270	0.196
28	210.4	193.4	17.0	486.4	216	200	780.4	206	0.276	0.256

TABLE II

FILM COEFFICIENTS OF CONDENSING PURE VAPORS

C TOLUENE

Run No.	Temperature, °F.			Condensate Gms. / Min.	Film Coefficient Btu / Hr. Ft.² °F.		Physical Property Group	Reynolds No.	h / ϕ	
	Vapor	Avg. Wall Temp.	ΔT	W	Experimental	Theoretical	ϕ	$\frac{4\Gamma}{\mu_f}$	Experimental	Theoretical
					FIRST SERIES					
1	230.4	174.2	56.2	1961.0	302	163	914.0	612	0.331	0.178
2	230.5	181.0	49.5	1658.8	290	171	916.0	531	0.317	0.187
3	230.4	182.8	47.6	1602.0	291	173	918.0	516	0.317	0.188
4	230.6	209.2	21.4	620.0	251	234	934.0	219	0.268	0.251
5	230.4	199.7	30.7	834.0	235	213	927.6	285	0.253	0.230
6	230.4	188.3	42.1	1370.0	282	182	922.0	449	0.306	0.197
7	230.4	202.7	27.7	782.2	244	217	930.2	270	0.263	0.234
8	230.4	210.5	19.9	588.0	256	238	934.8	208	0.274	0.255
9	230.7	208.8	21.9	624.0	247	234	934.0	220	0.264	0.250
10	230.6	208.6	22.0	640.0	252	232	933.7	225	0.270	0.248
11	230.6	199.2	31.4	837.2	231	213	928.5	286	0.249	0.230
12	230.6	193.8	36.8	1130.2	266	193	925.2	378	0.288	0.209
13	230.7	196.8	33.9	912.0	233	207	927.0	309	0.251	0.224
14	231.0	189.5	41.5	1334.0	278	183	923.0	440	0.302	0.198
15	230.4	210.5	19.9	597.0	260	237	935.0	211	0.278	0.254
16	230.6	197.2	33.4	900.0	233	208	927.0	305	0.252	0.225
17	230.6	206.3	24.3	684.4	244	227	932.5	239	0.267	0.243
18	230.6	193.6	37.0	1143.3	268	193	925.0	382	0.289	0.209
19	231.0	187.5	43.5	1443.2	287	179	921.7	472	0.312	0.194
20	231.0	187.1	43.9	1467.4	289	178	921.5	480	0.314	0.193
21	230.8	185.2	45.6	1555.0	295	175	920.4	505	0.321	0.190
					SECOND SERIES					
1	230.5	215.8	14.7	474.2	279	254	938.0	170	0.298	0.271
2	230.4	214.1	16.3	528.4	281	246	937.0	189	0.300	0.263
3	230.4	213.6	16.8	540.0	277	245	936.5	193	0.296	0.261
4	230.5	210.2	20.3	628.2	268	233	934.5	223	0.287	0.249
5	230.4	206.1	24.3	732.8	261	222	932.2	256	0.280	0.238
6	230.6	200.7	29.9	851.1	247	212	929.2	292	0.265	0.228
7	230.9	197.4	33.5	920.4	238	207	927.5	312	0.256	0.223
8	231.0	198.2	32.8	906.8	239	207	927.7	308	0.258	0.224
9	230.6	193.2	37.4	1212.9	282	189	925.0	405	0.304	0.204
10	231.0	191.8	39.2	1312.0	290	184	924.0	436	0.314	0.199
11	230.8	189.5	41.3	1409.7	296	180	922.7	465	0.320	0.195
12	230.8	186.0	44.8	1585.9	308	173	920.7	516	0.333	0.188
13	230.7	182.3	48.4	1784.9	319	167	918.6	574	0.348	0.182

TABLE II

FILM COEFFICIENTS OF CONDENSING PURE VAPORS

D n-HEXANE

Run No.	Temperature, °F.			Condensate Gms. / Min.	Film Coefficient Btu / Hr. Ft.^{2} °F.		Physical Property Group	Reynolds No.	h / φ	
	Vapor	Avg. Wall Temp.	ΔT	W	Experimental	Theoretical	φ	$\frac{4r}{\mu_f}$	Experimental	Theoretical
					FIRST SERIES					
1	156.7	147.6	9.1	283.0	248	251	829.8	125	0.299	0.302
2	157.0	147.8	9.2	287.1	249	249	829.8	127	0.300	0.300
3	156.8	141.0	15.8	452.8	229	215	829.8	196	0.276	0.260
4	156.9	141.0	15.9	440.9	221	217	828.0	191	0.267	0.262
5	156.9	137.4	19.5	540.9	222	204	826.6	232	0.268	0.246
6	156.8	137.2	19.6	546.3	222	203	828.3	234	0.268	0.245
7	157.0	134.0	23.0	610.4	212	196	827.8	258	0.256	0.237
8	157.0	129.3	27.7	696.9	200	189	827.0	290	0.242	0.228
9	157.0	129.2	27.8	700.0	201	188	827.0	291	0.243	0.228
10	156.9	124.0	32.9	796.7	193	181	826.1	325	0.234	0.219
11	157.1	123.9	33.2	810.7	195	180	826.1	331	0.236	0.218
12	157.0	119.2	37.8	1013.5	214	168	825.4	407	0.259	0.204
13	157.0	119.2	37.8	1028.1	217	167	825.4	413	0.263	0.203
14	157.0	115.0	42.0	1221.8	232	159	824.4	483	0.282	0.192
15	156.9	144.2	12.7	385.0	242	226	829.3	169	0.292	0.273
16	157.0	144.4	12.6	381.2	241	227	829.4	167	0.291	0.274
					SECOND SERIES					
1	156.8	153.8	3.0	165.5	440	298	830.8	75	0.530	0.359
2	156.9	152.5	4.4	210.6	382	276	830.6	95	0.460	0.332
3	156.9	150.4	6.5	268.3	329	254	830.3	120	0.397	0.306
4	156.7	150.5	6.2	257.3	331	258	830.3	115	0.399	0.311
5	156.8	144.2	12.6	422.0	268	220	829.3	185	0.324	0.265
6	156.9	144.2	12.7	429.0	270	219	829.3	188	0.325	0.264
7	156.9	144.2	12.7	440.0	276	217	829.3	193	0.333	0.261
8	156.9	140.8	16.1	511.2	253	207	828.8	221	0.306	0.250
9	157.0	136.8	20.2	607.5	241	196	828.2	259	0.290	0.237
10	157.0	136.8	20.2	608.8	240	196	828.2	260	0.290	0.237
11	157.0	129.7	27.3	734.9	215	185	827.1	306	0.260	0.223
12	157.0	131.8	25.2	730.0	231	185	827.4	306	0.279	0.224
13	156.9	127.2	29.7	811.6	218	180	826.6	335	0.264	0.217

coefficients as a function of temperature drop across the film (Figures 6 to 9), and (2) a plot of h/ϕ as a function of Reynolds number (Figures 10 to 13). The heat transfer coefficients were measured, for benzene only, at two different pressure ranges in the jacket surrounding the primary condenser. One set of data were taken when pressure of 2 to 4 inches of water was maintained in the jacket, and the other set was determined under a pressure of 1 to 4 inches Mercury. These determinations were carried out to investigate the effect of the vapor velocity on the film coefficient. The vapor velocities in the jacket and the auxiliary condenser varied between 0.45 to 0.64 ft/sec. and 2.5 to 6.5 ft/sec., respectively. The effects of these vapor velocities on the heat transfer coefficient for benzene are shown in Figures 6 and 10. It is noticeable on both diagrams that the data which were determined at a jacket pressure of less than four inches of water all fall on a single straight line. Those determined at a higher jacket pressure fall above this line, their deviation depending upon the jacket pressure. Investigation on benzene showed that maintaining a pressure of 2 to 4 inches of water in the jacket would be sufficient to keep the air and non-condensables out of the system by allowing a small quantity of vapor to enter the auxiliary condenser. The coefficients for toluene and n-hexane were re-measured subsequent to the time when it was noticed that the condensing bar was covered with a thin layer of deposit and was tarnished. The unit was dismantled and the bar was taken out and polished. The coefficients measured after cleaning the bar were 5 to 10 per cent higher than previous measurements made for toluene and n-hexane.

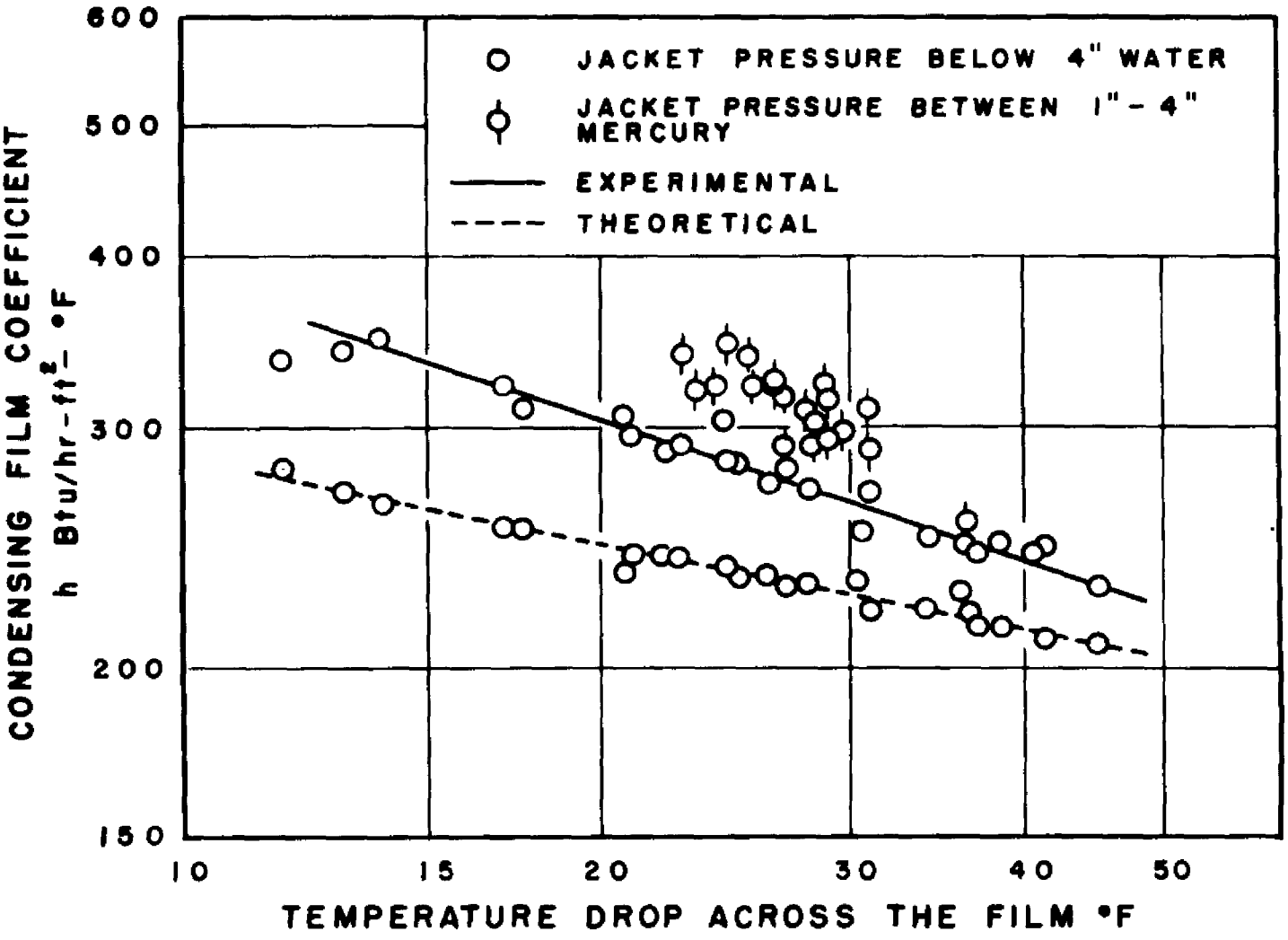


FIGURE 6. FILM COEFFICIENT OF HEAT TRANSFER FOR CONDENSING BENZENE VAPOR

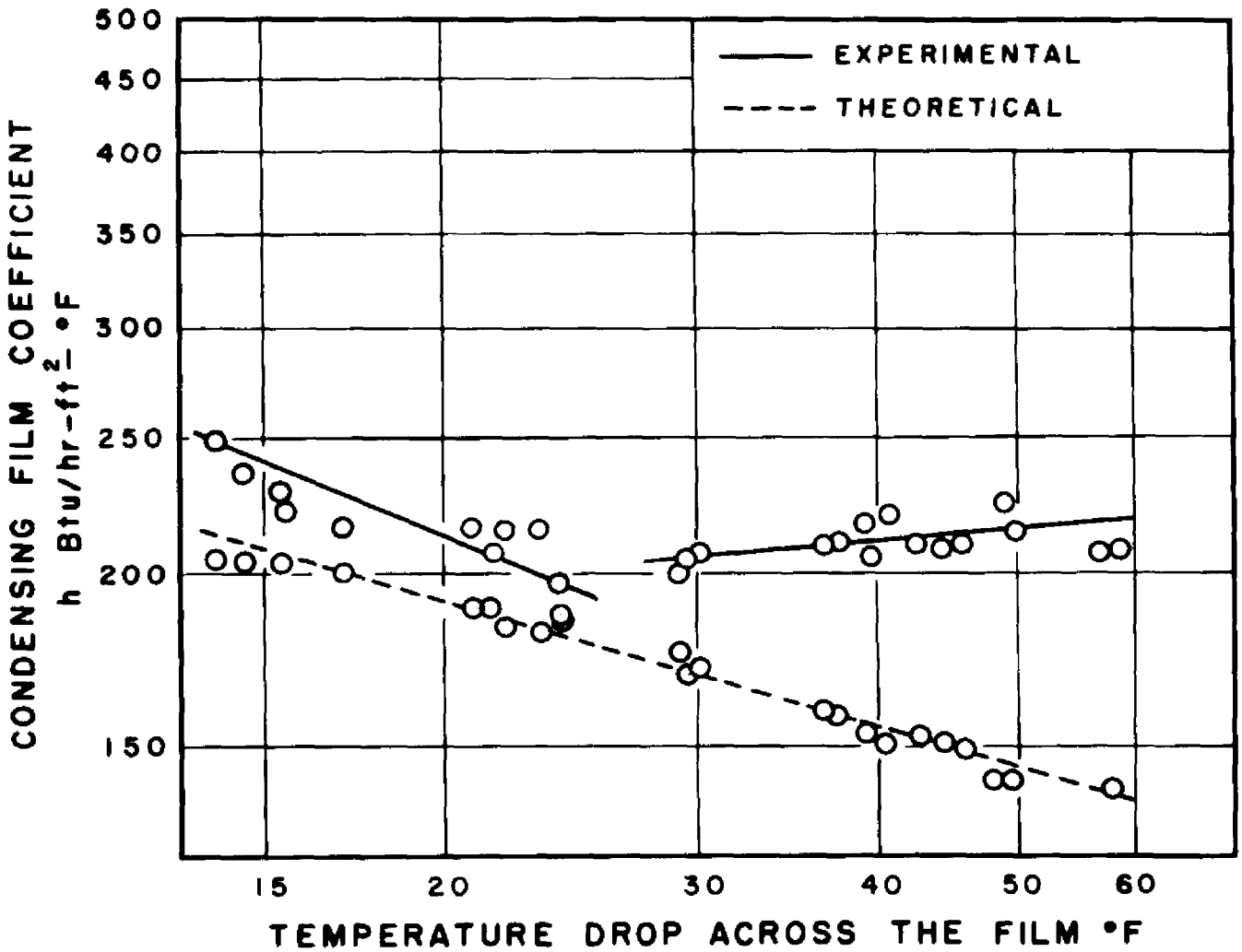


FIGURE 7. FILM COEFFICIENT OF HEAT TRANSFER FOR CONDENSING N-HEPTANE VAPOR

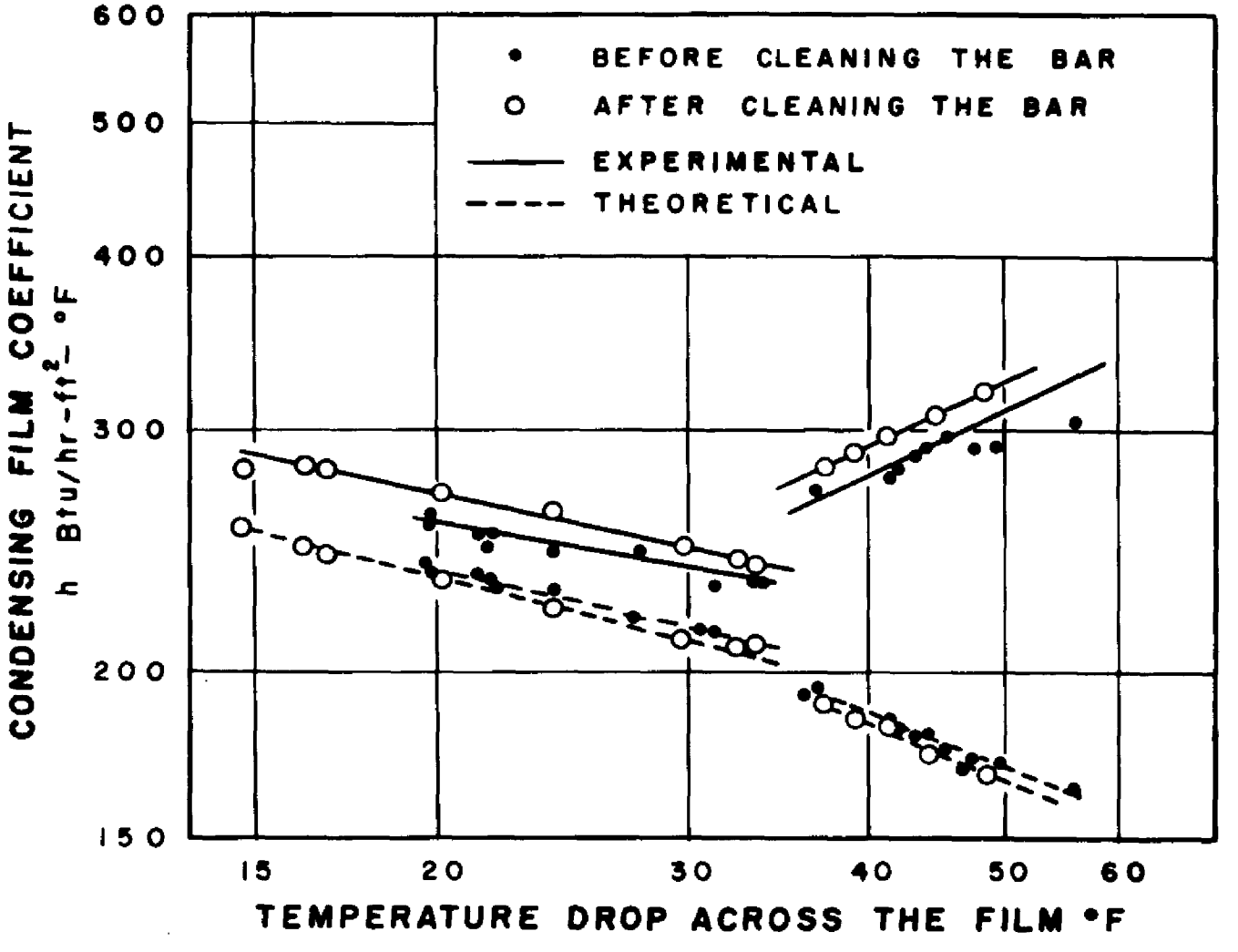


FIGURE 8. FILM COEFFICIENT OF HEAT TRANSFER FOR CONDENSING TOLUENE VAPOR

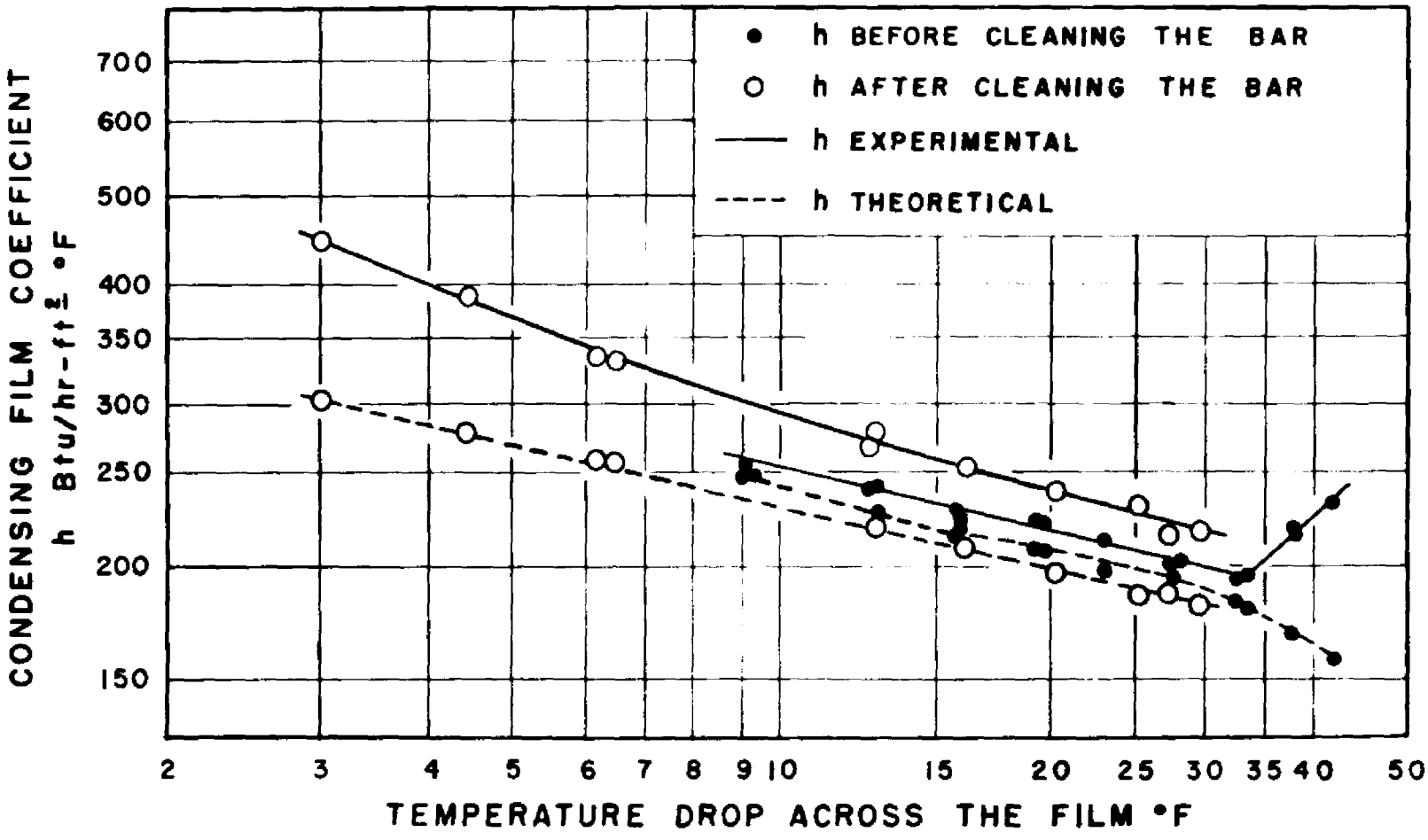


FIGURE 9. FILM COEFFICIENT OF HEAT TRANSFER OF CONDENSING N-HEXANE VAPOR

The reproducibility of the results was determined. Duplicate tests generally gave coefficients that differed by less than 2 to 3 per cent from the average of the two. The film coefficients of heat transfer for benzene also were determined by Todd (104) and Malone (59). Under the same operating conditions, the present results are about 40 per cent higher than those obtained by Malone. But they coincide exactly with those obtained by Todd (104) when the unit was first operated. The low coefficients obtained by Malone can be attributed to a progressive fouling of the surface over a long period, not removed by the in-place cleaning method he employed.

All of the heat transfer coefficients for benzene decrease with increase in temperature drop across the film, whereas for n-heptane, toluene, and n-hexane the decrease in heat transfer coefficient is up to a certain temperature drop only, after which coefficients rise with a further increase in temperature drop. The temperature difference where heat transfer coefficients start increasing is in the range of 25 to 35°F., differing from one hydrocarbon to another. This point will be discussed in more detail in a later paragraph.

The slopes of the lines of film coefficient versus temperature drop were calculated for the pure hydrocarbons and are presented below for comparison with the theoretical slope of -0.25. Here it should be pointed out that Nusselt derivation assumes that the physical properties of the condensate are constant throughout the layer, whereas the presence of a temperature gradient means there is some variation in these properties, even though small in many cases.

Pure Compound	Slope
Benzene	-0.30
Toluene	-0.23
n-Heptane	-0.35
n-Hexane	-0.24

The deviation from theoretical value is attributed to the possible existence of turbulence in the film, the change in film temperature and the corresponding change in the physical property group used in Equation 8.

The theoretical coefficients for each hydrocarbon are calculated by Nusselt's equation for the observed condensation rate and film temperature. These coefficients are plotted for comparison with the experimental data. Generally, the experimental coefficients are 20 to 30 per cent higher than the theoretical values. Part of this difference could be explained by the presence of impurities in the hydrocarbons used. This effect will be more pronounced when considering the fact that the physical property data used in calculating the theoretical coefficients, which were taken from literature, are for pure compounds and do not represent the actual properties of the commercial-grade hydrocarbons used here.

Nusselt's theory is based upon the assumption of streamline flow of the condensate. However, turbulence will occur when the film thickness and velocity exceed critical values. The present work shows that the experimental heat transfer coefficients increase, in comparison with the corresponding theoretical values, as the rate of heat transfer increases. The presence of turbulent flow in the condensate at high rate of heat transfer would explain this qualitatively.

McAdams (62) summarized the experimental data for film type condensation of steam and 18 organic vapors, outside single horizontal tubes. The data were compared with the values predicted with the theoretical Nusselt's equation for laminar flow of condensate, neglecting any effect of vapor velocity. The measured values of the heat transfer coefficient ran from 36 per cent below to 70 per cent above those predicted from the same values of temperature drop across the film. The average of the ratio of measured to predicted coefficients was 1.23 for steam and 0.94 for the organic vapors. The ratios of the average of the experimental data to the predicted theoretical values for the present work and those published in literature are tabulated as follows:

	Ratio of experimental to theoretical coefficients	
Hydrocarbon	Present Investigation	Others reported by McAdams
Benzene	1.18	0.8 - 1.22
Toluene	1.32	0.76- 1.04
n-Heptane	1.25	-
n-Hexane	1.28	-

The ratios reported from McAdams cover the data of various investigators and are for a wide range of condenser sizes and operating conditions.

The results for pure hydrocarbons are also plotted in terms of h/ϕ as a function of Reynolds number, $4\Gamma/\mu_f$, of the condensate. These are shown on Figures 10 through 13 for the individual hydrocarbons and on Figure 14 for all of the four hydrocarbons tested. Employing the least square method, the slope and the intercept of the experimental line were

determined to be -0.463 and 3.6 as compared to the theoretical values of -1/3 and 1.51. The theoretical results based on the observed temperature drops across the film and condensate rates are also plotted for comparison. The accuracy of these results is the same as those of heat transfer coefficient and temperature difference across the film. However, they show that the Reynolds number of the condensate never exceeded 600.

The accepted critical value of condensate Reynolds number is 4200; below that the condensate is supposedly in streamline flow. Therefore, the condensate with Reynolds number below 600 should be in streamline flow. This may be true in some instances, but it was not so in cases of n-heptane, toluene, and n-hexane: for these at a Reynolds number of about 300, the heat transfer coefficient became higher (contrary to the theory for streamline flow) as the Reynolds number increased. Figures 11, 12, and 13 show that the condensate followed the theory behavior only up to Reynolds number of about 300. This indicates that the accepted critical value of Reynolds number of 4200, is not an accurate criteria of the turbulence, and that the behavior indicates that turbulence starts at a much lower range of Reynolds number. These figures indicate that the condensate layer, on the outside of a horizontal tube, is essentially in viscous flow up to Reynolds number of about 300, but above 300 it breaks into turbulent flow. There have been no experimental data reported in the literature to show at what Reynolds number the turbulence starts when condensing pure vapors on the outside of a horizontal tube.

Carpenter and Colburn (17) made a study of the effect of vapor velocity on condensation inside long vertical tubes. Comparison of

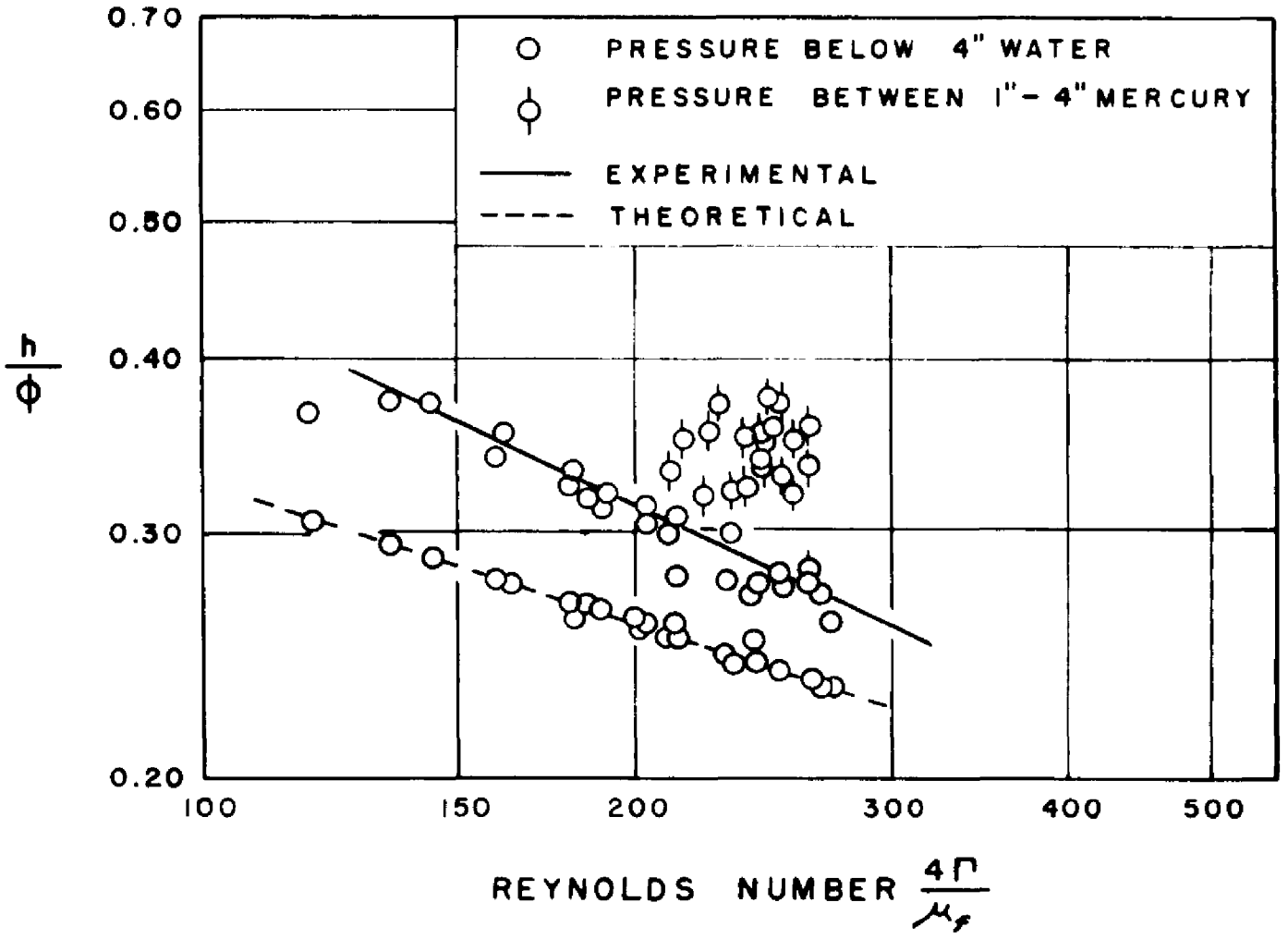


FIGURE 10. VARIATION OF h/ϕ WITH REYNOLDS NUMBER FOR CONDENSING BENZENE VAPOR

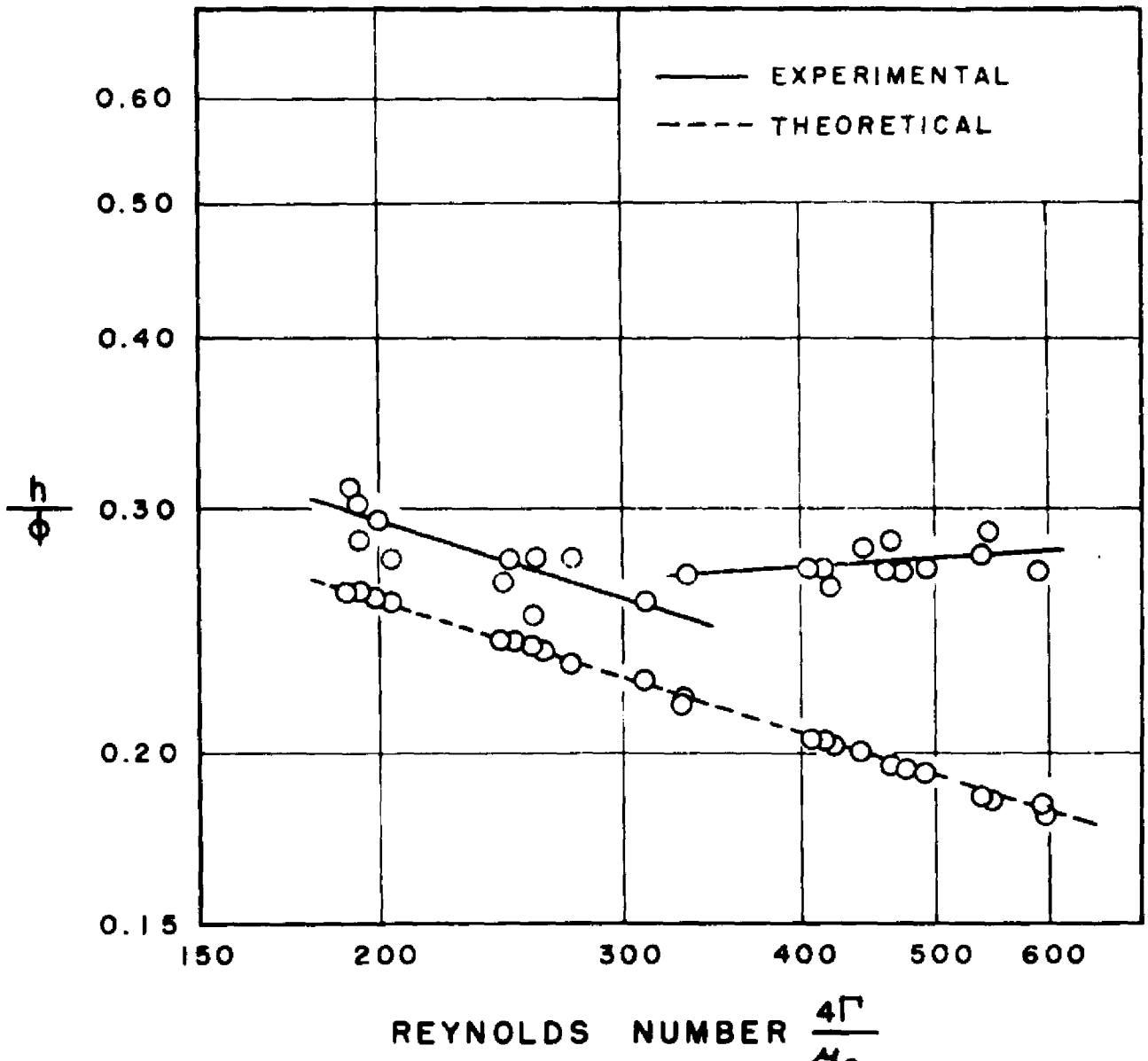


FIGURE 11. VARIATION OF h/ϕ WITH REYNOLDS NUMBER FOR CONDENSING N-HEPTANE VAPOR

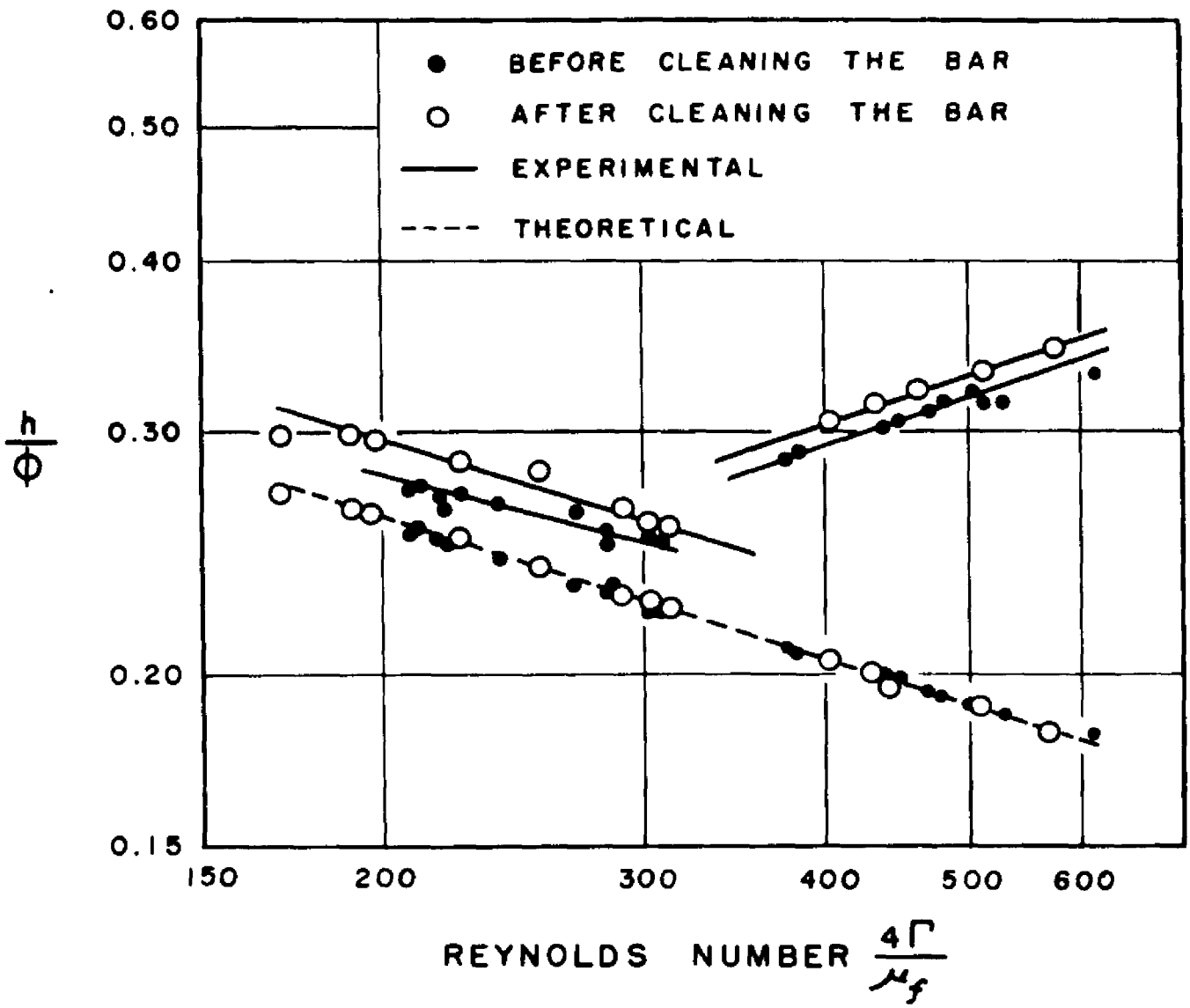


FIGURE 12. VARIATION OF h/ϕ WITH REYNOLDS NUMBER FOR CONDENSING TOLUENE VAPOR

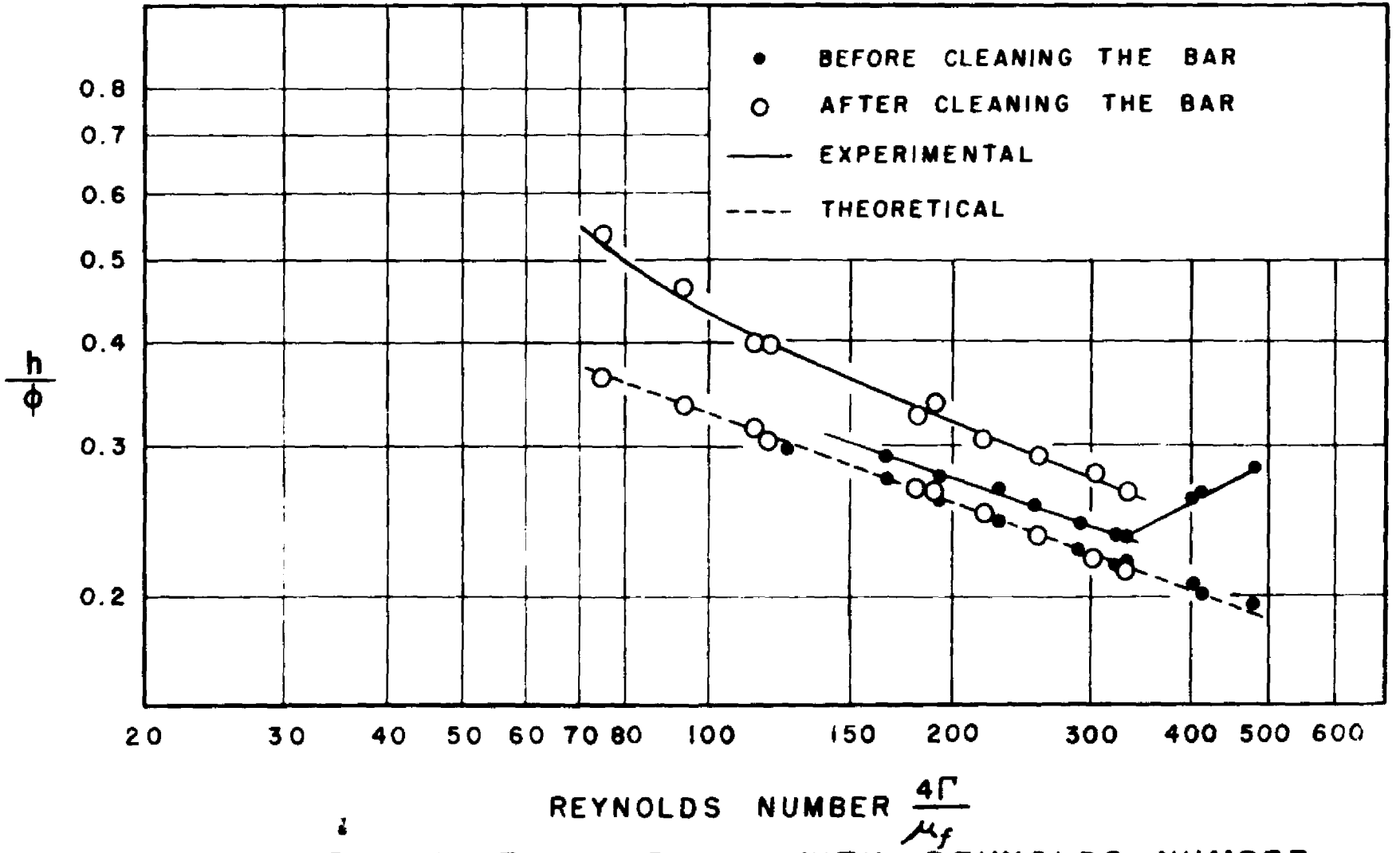


FIGURE 13. VARIATION OF h/ϕ WITH REYNOLDS NUMBER FOR CONDENSING N-HEXANE VAPOR

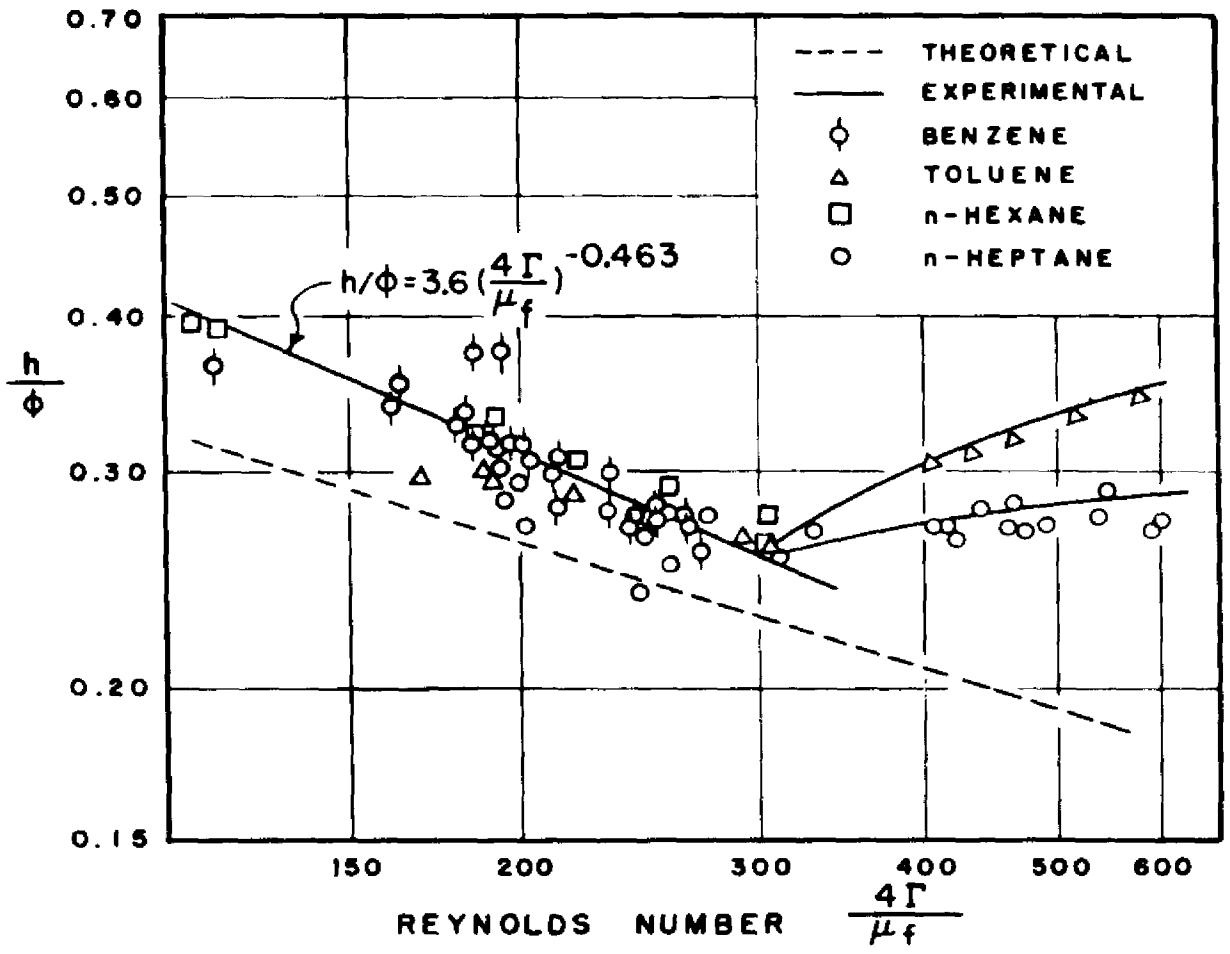


FIGURE 14. FILM-TYPE CONDENSATION OF SINGLE VAPORS ON HORIZONTAL TUBE

their results with those predicted by a modified equation developed by Nusselt, which includes the effect of vapor friction on the thickness of the viscous condensate layer, showed that their values of condensing film coefficient were higher than predictions, in some cases several-fold higher. In his modification, Nusselt assumed that the effect of vapor friction was limited to consideration of a condensate layer in viscous flow. The effect of vapor friction was believed to increase the downward force on the film, and thus decrease the film thickness. Caprenter-Colburn hypothesis was that "in the presence of a high frictional force from the vapor on its outer surface, the condensate layer would become turbulent at a much lower value of Reynolds number than found when vapor friction was negligible". Based upon their experimental data, it was concluded by the authors that when vapors condense in the presence of high vapor velocities, turbulence is induced in the condensate layer, thereby lowering the effective film thickness of the condensate layer. It was also concluded that whereas in the absence of vapor velocity when the condensate layer is essentially in laminar flow up to a Reynolds number of about 2000, in the presence of high vapor velocity of about one to three hundred feet per second downward inside of a vertical condenser, the layers became turbulent at a Reynolds number of about 240.

Based upon the foregoing, it is reasonable to deduce that in the present study, too, a turbulence was present in the condensate layer, which resulted experimental coefficients to be 20 to 30 per cent higher than theoretical values, even though the maximum vapor velocity in the

jacket was less than one foot per second. For calculating the vapor velocity, it was assumed that one-half of total vapor entering the primary condenser flowed through each section of the annulus between the jacket and the condensing bar.

The effect of non-linear temperature distribution and the cross-flow within the film on condensing heat transfer coefficient as proposed by Bromley (10) and Rohsenow (81) was studied for the four hydrocarbons. Except for Test Numbers 5, 6, 19, and 20 for n-heptane, where the magnitude of the liquid sub-cooling or more precisely the value of $C_{pf}\ \Delta T/\lambda$ were 0.214, 0.209, 0.250, and 0.252, respectively, for all other tests it was less than 0.2. Accordingly, the latent heat of condensation was modified as proposed by Rohsenow, Equation 12, Chapter II, and the modified latent heat term was used to correct the calculated theoretical heat transfer coefficients. These calculations are summarized in Table III-A through D. It will be noticed that the Rohsenow correction does not increase the condensing heat transfer coefficient significantly. The maximum increase is 3.1 Btu/Hr.Ft.°F for Test No. 20 of n-heptane, which is an increase of only 2.2 per cent. The Rohsenow correction factor was not determined for all of the tests made for the four hydrocarbons. However, a sufficient number of tests were included to cover the maximum range of the liquid sub-cooling term, $C_{pf}\ \Delta T/\lambda$.

II - Binary Vapors

Experiments with the pure hydrocarbons indicated that maintaining a slight positive pressure (2 to 4 inches of water) in the jacket was sufficient to insure the absence of non-condensables in the vapor

TABLE III

ROHSENOW'S (81) CORRECTION TO THE NUSSELT'S EQUATION OF HEAT TRANSFER COEFFICIENT FOR CONDENSING PURE VAPORS

$$\lambda' = \lambda + 3/8\ (C_{pf}\ \Delta T)$$

A Benzene λ = 169.34 Btu/Lb.

Run No.	ΔT °F.	C_{pf} Btu/Lb.°F.	$\frac{C_{pf}\ \Delta T}{\lambda}$	λ' Btu/Lb.	h'_m Btu/Hr. Ft.^{2} °F.	h_m Btu/Hr. Ft.^{2} °F.	$h'_m - h_m$ Btu/Hr. Ft.^{2} °F.
2	36.8	0.4470	0.097	175.5	218.9	217	1.9
3	38.1	0.4465	0.100	175.7	216.0	214	2.0
5	41.0	0.4450	0.107	176.2	210.0	208	2.0
7	24.5	0.4507	0.065	173.5	237.4	236	1.4
9	20.7	0.4515	0.050	172.5	238.1	237	1.1
11	21.1	0.4515	0.056	172.9	244.3	243	1.3
13	27.0	0.4500	0.072	173.9	229.4	228	1.4
14	24.9	0.4507	0.066	173.5	233.4	232	1.4
15	28.1	0.4496	0.074	174.1	229.6	228	1.6
39	22.2	0.4517	0.059	173.1	242.3	241	1.3
40	22.6	0.4517	0.060	173.2	241.4	240	1.4
41	17.6	0.4530	0.047	172.3	254.0	253	1.0
42	17.1	0.4532	0.046	172.3	254.0	253	1.0
43	13.9	0.4542	0.037	171.7	264.9	264	0.9
44	13.1	0.4545	0.035	171.6	270.9	270	0.9
45	31.0	0.4490	0.082	174.5	220.7	219	1.7
46	30.6	0.4490	0.081	174.5	233.7	232	1.7
47	41.2	0.4458	0.108	176.2	211.0	209	2.0
48	45.0	0.4452	0.118	176.8	209.2	207	2.2

TABLE III

ROHSENOW'S (81) CORRECTION TO THE NUSSELT'S EQUATION OF HEAT TRANSFER COEFFICIENT FOR CONDENSING PURE VAPORS

$$\lambda' = \lambda + 3/8\ (C_{pf}\ \Delta T)$$

B n-Heptane λ = 135.99 Btu/Lb.

Run No.	ΔT °F.	C_{pf} Btu/Lb.°F.	$\frac{C_{pf}\ \Delta T}{\lambda}$	λ' Btu/Lb.	h'_m Btu/Hr. Ft.^{2} °F.	h_m Btu/Hr. Ft.^{2} °F.	$h'_m - h_m$ Btu/Hr. Ft.^{2} °F.
4	36.8	0.5940	0.160	144.2	160.3	158	2.3
5	49.8	0.5842	0.214	147.2	144.8	142	2.8
6	48.4	0.5890	0.209	146.6	144.6	142	2.6
7	40.4	0.5915	0.178	144.9	152.4	150	2.4
8	29.5	0.5975	0.129	142.6	171.0	169	2.0
9	23.3	0.6000	0.103	141.2	182.7	181	1.7
10	29.0	0.5980	0.127	142.5	177.0	175	2.0
11	30.0	0.5970	0.132	142.7	171.0	169	2.0
12	21.8	0.6010	0.096	140.9	186.6	185	1.6
13	22.1	0.6008	0.097	141.0	185.6	184	1.6
14	15.6	0.6037	0.069	139.5	204.3	203	1.3
15	15.5	0.6037	0.068	139.5	203.2	202	1.2
16	14.6	0.6042	0.065	139.3	205.2	204	1.2
17	39.1	0.5930	0.170	144.7	156.4	154	2.4
18	45.7	0.5900	0.198	146.1	150.7	148	2.7
19	58.0	0.5850	0.250	148.7	141.1	139	2.1
20	58.8	0.5825	0.252	148.8	141.1	138	3.1
21	13.7	0.6025	0.060	139.1	205.8	204	1.8
26	44.2	0.5905	0.191	145.8	152.6	150	2.6
27	42.8	0.5906	0.185	145.4	153.6	151	2.6

TABLE III

ROHSENOW'S (81) CORRECTION TO THE NUSSELT'S EQUATION OF HEAT TRANSFER COEFFICIENT FOR CONDENSING PURE VAPORS

$$\lambda' = \lambda + 3/8\ (C_{pf}\ \Delta T)$$

C Toluene (Second Series) λ = 156.2 Btu/Lb.

Run No.	ΔT °F.	C_{pf} Btu/Lb.°F.	$\frac{C_{pf}\ \Delta T}{\lambda}$	λ' Btu/Lb.	h'_m	h_m	$h'_m - h_m$
					Btu/Hr. Ft.^{2} °F.		
1	14.7	0.4750	0.0446	158.8	255.0	254	1.0
2	16.3	0.4745	0.0495	159.1	247.1	246	1.1
3	16.8	0.4742	0.0512	159.2	246.2	245	1.2
4	20.3	0.4735	0.0615	159.8	234.3	233	1.3
5	24.3	0.4720	0.0732	160.5	223.5	222	1.5
6	29.9	0.4710	0.0900	161.5	213.8	212	1.8
7	33.5	0.4697	0.1005	162.1	208.9	207	1.9
8	32.8	0.4700	0.0985	162.0	208.9	207	1.9
9	37.4	0.4685	0.1120	162.8	190.9	189	1.9
10	39.2	0.4680	0.1172	163.1	186.0	184	2.0
11	41.3	0.4670	0.1232	165.4	182.0	180	2.0
12	44.8	0.4660	0.1335	164.0	175.1	173	2.1
13	48.4	0.4650	0.1440	164.6	169.2	167	2.2

TABLE III

ROHSENOW'S (81) CORRECTION TO THE NUSSELT'S EQUATION OF HEAT TRANSFER COEFFICIENT FOR CONDENSING PURE VAPORS

$$\lambda' = \lambda + 3/8\ (C_{pf}\ \Delta T)$$

D n-Hexane λ = 143.96 Btu/Lb.

Run No.	ΔT °F.	C_{pf} Btu/Lb.°F.	$\frac{C_{pf}\ \Delta T}{\lambda}$	λ' Btu/Lb.	h'_m	h_m	$h'_m - h_m$
					Btu/Hr. Ft.^{2} °F.		
1st Series							
11	33.2	0.5770	0.1330	151.4	182.4	180	2.2
12	37.8	0.5745	0.1530	152.4	170.4	168	2.4
14	42.0	0.5725	0.1670	153.0	161.4	159	2.4
2nd Series							
1	3.0	0.5940	0.0124	144.6	298.3	298	0.3
3	6.5	0.5905	0.0267	145.4	254.6	254	0.6
6	12.7	0.5880	0.0518	146.8	220.0	219	1.0
8	16.1	0.5855	0.0655	147.5	208.3	207	1.3
10	20.2	0.5840	0.0818	148.4	197.5	196	1.5
11	27.3	0.5800	0.1100	149.9	186.9	185	1.9
12	25.2	0.5810	0.1017	149.5	186.7	185	1.7
13	29.7	0.5790	0.1195	150.4	182.0	180	2.0

in the jacket surrounding the condenser; by allowing a slight amount of vapor to flow to the auxiliary condenser. For binary vapors maintenance of this positive pressure caused the entering vapor temperature, when equilibrium was established, to be 0.3 to 2°F. above that of the dew point of the condensate at one atmosphere.

In condensing the binary vapors the experiments were carried out in such a manner as to have almost total condensation of vapor and a condensate of constant composition. Material balance data on the unit which are summarized in Table IV-A, B and C show that, except for a very few cases, composition of condensate from the primary condenser differed by less than one mole per cent from the composition of the vapor from the reboiler. Quite frequently, for better than 85 per cent of the tests made on all binary mixtures, the difference between the two composition was less than 0.5 mole per cent. The analyses of samples from the auxiliary condenser showed it to be consistently richer in the more volatile component than the condensate from the primary condenser obtained at the same time. The McCabe-Thiele step-by-step method was used to find the number of ideal plates between the compositions of the condensate from the primary condenser and the condensate from the auxiliary condenser. It was found that equivalent to a fraction of a theoretical plate existed in the apparatus, for all of the tests performed. As it is pointed out in Appendix C, Section II-B, the Murphree plate efficiency equation may be used to express the degree of enrichment of the residual vapors. Data in Table VII-A, B, and C show that the degree of enrichment of the residual vapor, or the approach of the residual vapor to equilibrium with the primary

TABLE IV

COMPOSITION OF VAPOR TO THE PRIMARY CONDENSER

A BENZENE - n-HEPTANE SYSTEM

Run No.	Benzene in Primary Condensate		Benzene in Auxiliary Condensate		Condensate Gms. / Min.		W/Wa	Benzene in Vapor to Primary Condenser		Mole % Difference Between Vapor to and Condensate from the Primary Condenser
	Mole % x_A	Wt. % X_A	Mole % x_{Aa}	Wt. % Y_{Aa}	Primary W	Auxiliary Wa	R	Wt. % Y_{Al}	Mole % y_{Al}	$y_{Al} - x_A$
			A 82.5 MOLE PER CENT BENZENE IN THE PRIMARY CONDENSATE							
1	82.6	78.72	85.2	81.77	276.7	13.0	21.28	78.85	82.72	+0.12
2	82.6	78.72	85.2	81.77	388.9	12.9	30.22	78.82	82.69	+0.09
3	82.6	78.72	85.2	81.77	494.5	12.9	38.21	78.80	82.67	+0.07
4	82.4	78.49	85.3	81.89	601.8	14.3	42.08	78.57	82.47	+0.07
5	82.4	78.49	85.2	81.77	639.7	11.8	54.21	78.55	82.46	+0.06
6	82.5	78.60	85.3	81.89	762.5	11.0	69.32	78.65	82.54	+0.04
7	82.5	78.60	84.8	81.30	343.5	13.2	26.02	78.70	82.59	+0.09
8	82.6	78.72	84.8	81.30	446.6	13.5	33.08	78.80	82.67	+0.07
9	82.4	78.49	84.8	81.30	546.0	13.5	40.44	78.56	82.47	+0.04
10	82.4	78.49	85.2	81.77	842.5	11.5	73.26	78.53	82.44	+0.04
Average	82.5	78.60	85.1					78.68	82.57	+0.07
			B 70.2 MOLE PER CENT BENZENE IN THE PRIMARY CONDENSATE							
1	70.0	64.52	75.7	70.83	248.6	12.0	20.72	64.81	70.26	+0.26
2	70.2	64.74	76.0	71.17	232.7	12.6	18.47	65.07	70.50	+0.30
3	69.9	64.47	76.0	71.17	304.5	13.5	22.55	64.75	70.21	+0.31
4	70.2	64.74	75.8	70.94	449.4	14.9	30.16	64.93	70.37	+0.17
5	70.2	64.74	76.2	71.39	444.6	15.2	29.25	64.95	70.39	+0.19
6	70.2	64.74	75.8	70.94	576.8	15.3	37.70	64.90	70.35	+0.15
7	70.2	64.74	76.2	71.39	625.8	15.8	39.60	64.90	70.35	+0.15
8	70.3	64.85	76.4	71.61	615.6	16.0	38.47	65.02	70.46	+0.16
9	70.2	64.74	75.7	70.83	711.0	18.0	39.50	64.89	70.34	+0.14
10	70.3	64.85	75.7	70.83	708.3	18.0	39.35	65.00	70.44	+0.14
11	70.3	64.85	75.6	70.72	463.2	16.0	28.95	65.04	70.48	+0.18
12	70.1	64.63	75.6	70.72	385.5	15.1	25.50	64.86	70.31	+0.21
Average	70.2	64.72	75.9	71.05				64.93	70.37	+0.17
			C 54.6 MOLE PER CENT BENZENE IN THE PRIMARY CONDENSATE							
1	54.8	48.58	64.0	58.08	245.0	18.0	13.61	49.27	55.44	+0.64
2	54.8	48.58	64.0	58.08	336.5	17.2	19.56	49.04	55.24	+0.44
3	54.5	48.28	63.4	57.45	441.3	18.0	24.52	48.64	54.85	+0.35
4	54.9	48.69	64.2	58.29	552.5	15.5	35.64	48.95	55.15	+0.25
5	54.6	48.38	64.0	58.08	578.3	15.5	37.31	48.62	54.83	+0.23
6	54.4	48.18	63.9	57.98	637.0	17.0	37.47	48.43	54.65	+0.25
7	54.4	48.18	63.7	57.77	718.5	20.0	35.92	48.43	54.65	+0.25
8	54.5	48.28	64.1	58.19	732.9	17.5	41.88	48.51	54.73	+0.23
9	54.5	48.28	63.8	57.87	330.2	18.1	18.24	48.78	54.99	+0.49
10	54.6	48.38	64.0	58.08	580.2	16.0	36.26	48.64	54.85	+0.25
Average	54.6	48.38	63.9	57.99				48.73	54.94	+0.34
			D 32.6 MOLE PER CENT BENZENE IN THE PRIMARY CONDENSATE							
1	32.6	27.40	38.5	32.80	257.5	20.2	12.75	27.80	33.06	+0.46
2	32.3	27.12	38.3	32.60	379.2	18.1	20.95	27.35	32.56	+0.26
3	32.8	27.74	39.8	34.06	317.3	22.0	14.42	28.10	33.39	+0.59
4	32.8	27.74	40.2	34.37	583.0	20.2	28.86	27.92	33.20	+0.40
5	32.8	27.74	40.0	34.25	653.3	16.5	39.59	27.86	33.13	+0.33
6	32.6	27.43	40.2	34.37	753.0	21.7	34.70	27.59	32.83	+0.23
7	32.3	27.14	40.0	34.25	524.8	19.8	26.50	27.36	32.58	+0.28
8	32.5	27.28	39.8	34.06	449.4	20.0	22.47	27.49	32.72	+0.22
9	32.6	27.43	39.9	34.17	425.9	18.2	23.40	27.67	32.92	+0.32
10	32.8	27.74	40.2	34.37	293.3	27.0	13.30	28.15	33.46	+0.65
Average	32.6	27.47	39.7	33.93				27.73	32.98	+0.38

TABLE IV

COMPOSITION OF VAPOR TO THE PRIMARY CONDENSER

B n-HEPTANE - TOLUENE SYSTEM

Run No.	n-Heptane in Primary Condensate		n-Heptane in Auxiliary Condensate		Condensate Cms. / Min.		W/W₂	n-Heptane in Vapor to Primary Condenser		Mole % Difference Between Vapor to and Condensate from the Primary Condenser
	Mole % x_A	Wt. % X_A	Mole % X_{A_2}	Wt. % Y_{A_2}	Primary W	Auxiliary W_2	R	Wt. % Y_{A1}	Mole % y_{A1}	$y_{A1}-x_A$
			A 85.6 MOLE PER CENT n-HEPTANE IN THE PRIMARY CONDENSATE							
1	85.6	86.61	86.7	87.64	317.2	36.0	8.81	86.70	85.70	+0.10
2	85.6	86.61	86.6	87.55	357.1	34.0	10.50	86.69	85.70	+0.10
3	85.5	86.51	86.7	87.64	406.7	38.0	10.70	86.62	85.62	+0.12
4	85.4	86.41	86.7	87.64	403.0	40.0	10.07	86.54	85.53	+0.13
5	85.7	86.70	86.7	87.64	385.0	36.0	10.69	86.79	85.78	+0.08
6	85.6	86.61	86.7	87.64	483.0	34.0	14.20	86.69	85.70	+0.10
7	85.6	86.61	86.7	87.64	530.1	33.0	16.06	86.67	85.66	+0.06
8	85.6	86.61	86.7	87.64	550.5	36.0	15.29	86.67	85.66	+0.06
9	85.6	86.61	86.6	87.55	594.0	31.1	19.10	86.65	85.64	+0.04
10	85.7	86.70	86.6	87.55	625.0	34.0	18.38	86.69	85.70	+0.10
11	85.7	86.70	86.6	87.55	622.5	37.0	16.82	86.74	85.75	+0.05
12	85.7	86.70	86.6	87.55	759.3	38.0	20.93	86.73	85.72	+0.02
13	85.6	86.61	86.7	87.64	733.4	43.0	17.06	86.67	85.66	+0.06
14	85.6	86.61	86.7	87.64	852.6	45.0	18.95	86.66	85.66	+0.06
Average	85.6	86.61	86.7	87.60				86.68	85.68	+0.08
			B 63.9 MOLE PER CENT n-HEPTANE IN THE PRIMARY CONDENSATE							
1	64.1	66.01	67.7	69.52	421.2	36.0	11.70	66.28	64.40	+0.30
2	64.1	66.01	67.8	69.61	523.0	32.0	16.34	66.22	64.32	+0.22
3	64.1	66.01	67.7	69.56	521.6	38.0	13.72	66.25	64.34	+0.24
4	63.9	65.82	67.8	69.61	654.4	38.1	17.18	66.02	64.12	+0.22
5	63.9	65.82	67.7	69.52	727.5	38.2	19.04	66.00	64.10	+0.20
6	63.9	65.87	67.7	69.55	727.5	38.0	19.14	66.05	64.13	+0.23
7	63.7	65.63	67.8	69.61	861.6	41.2	20.91	65.81	63.89	+0.19
8	63.8	65.72	67.8	69.61	569.4	39.0	14.60	65.97	64.06	+0.26
9	63.8	65.72	67.8	69.61	990.2	41.0	24.15	65.87	63.97	+0.17
10	63.8	65.72	67.8	69.61	1113.0	43.0	25.88	65.86	63.96	+0.16
Average	63.9	65.83	67.8	69.58				66.03	66.13	+0.23
			C 50.1 MOLE PER CENT n-HEPTANE IN THE PRIMARY CONDENSATE							
1	50.4	52.50	53.2	55.29	328.1	37.6	8.72	52.73	50.63	+0.23
2	50.0	52.10	53.2	55.29	425.8	35.5	11.99	52.36	50.26	+0.26
3	50.0	52.10	53.2	55.29	381.0	36.6	10.41	52.38	50.28	+0.28
4	50.4	52.50	53.1	55.19	518.0	38.2	13.56	52.68	50.58	+0.18
5	50.4	52.50	53.1	55.09	488.5	37.0	13.20	52.68	50.58	+0.18
6	50.0	52.10	52.8	54.89	665.7	38.0	17.52	52.25	50.15	+0.15
7	50.0	52.10	52.9	54.99	598.8	38.0	15.76	52.27	50.17	+0.17
8	50.0	52.10	53.0	55.09	792.8	39.0	20.33	52.24	50.14	+0.14
9	50.3	52.40	53.1	55.19	800.0	41.0	19.51	52.53	50.43	+0.13
10	49.9	52.00	53.2	55.29	1084.0	43.0	25.21	52.12	50.02	+0.12
Average	50.1	52.24	53.0	55.16				52.47	50.32	+0.22
			D 31.0 MOLE PER CENT n-HEPTANE IN THE PRIMARY CONDENSATE							
1	31.2	33.04	35.9	37.86	477.2	34.0	14.03	33.36	31.53	+0.33
2	31.2	33.04	35.8	37.76	532.7	40.5	13.15	33.37	31.52	+0.32
3	31.0	32.83	35.7	37.66	578.0	38.4	15.05	33.17	31.33	+0.33
4	30.8	32.62	36.0	37.96	689.7	40.2	17.15	32.91	31.08	+0.28
5	31.2	33.04	35.8	37.76	689.7	38.0	18.15	33.28	31.43	+0.23
6	31.3	33.14	35.8	37.76	782.8	39.0	20.07	33.36	31.53	+0.23
7	30.0	31.80	35.8	37.76	856.0	41.0	20.88	32.07	30.26	+0.26
8	31.4	33.24	35.8	37.76	1150.0	43.0	26.74	33.40	31.55	+0.15
9	30.5	32.32	35.8	37.76	1551.4	46.0	33.72	32.48	30.66	+0.16
10	30.9	32.72	35.8	37.76	1450.0	39.0	37.18	32.85	31.02	+0.12
Average	31.0	32.78	35.8	37.78				33.02	31.19	+0.19

TABLE IV

COMPOSITION OF VAPOR TO THE PRIMARY CONDENSER

C n-HEXANE - TOLUENE SYSTEM

Run No.	n-Hexane in Primary Condensate		n-Hexane in Auxiliary Condensate		Condensate Cms. / Min.		W/Ws	n-Hexane in Vapor to Primary Condenser		Mole % Difference Between Vapor to and Condensate from the Primary Condenser
	Mole %	Wt. %	Mole %	Wt. %	Primary	Auxiliary		Wt. %	Mole %	
	x_A	X_A	x_{As}	Y_{As}	W	Ws	R	Y_{Al}	y_{Al}	$y_{Al}-x_A$
			A 80.0 MOLE PER CENT n-HEXANE IN THE PRIMARY CONDENSATE							
1	79.9	78.82	86.6	85.81	848.9	43.2	19.67	79.16	80.23	+0.33
2	80.4	79.33	86.0	85.18	865.0	39.2	22.06	79.58	80.64	+0.24
3	79.8	78.71	86.2	85.39	682.6	38.6	17.68	79.07	80.15	+0.35
4	79.7	78.60	85.7	84.87	616.5	29.2	21.11	78.88	79.96	+0.26
5	80.3	79.23	85.2	84.34	563.7	37.3	15.11	79.54	80.61	+0.31
6	80.3	79.23	84.9	84.03	531.1	32.0	16.59	79.51	80.56	+0.26
7	80.3	79.23	84.8	83.92	501.8	37.5	13.38	79.20	80.58	+0.28
8	79.9	78.82	86.2	85.39	340.0	35.6	8.58	792.0	80.27	+0.37
9	79.6	78.50	86.1	85.29	299.6	41.5	7.22	79.32	80.40	+0.80
10	80.1	79.02	86.0	85.18	388.5	36.6	10.61	79.55	80.62	+0.52
Average	80.0	78.95	85.8	84.94				79.33	80.40	+0.40
			B 70.2 MOLE PER CENT n-HEXANE IN THE PRIMARY CONDENSATE							
1	70.2	68.80	78.3	77.15	804.7	51.2	15.72	69.30	70.68	+0.48
2	70.2	68.80	78.6	77.46	756.7	53.0	14.28	69.37	70.75	+0.55
3	70.1	68.69	78.0	76.84	680.0	46.0	14.78	69.20	70.58	+0.48
4	69.9	68.49	79.3	78.19	612.8	59.1	10.37	69.34	70.71	+0.81
5	70.6	69.21	79.0	77.88	526.1	46.0	11.43	69.90	71.22	+0.62
6	70.4	69.00	78.7	77.57	421.1	40.0	10.53	69.74	71.07	+0.67
7	70.5	69.10	78.3	77.15	282.4	47.0	6.00	70.25	71.55	+1.05
8	70.1	68.69	78.6	77.46	251.0	32.0	7.84	69.68	71.04	+0.94
9	70.2	68.80	79.5	78.40	464.9	31.0	15.00	69.40	70.77	+0.57
Average	70.2	68.84	78.7	77.57				69.57	70.93	+0.73
			C 50.2 MOLE PER CENT n-HEXANE IN THE PRIMARY CONDENSATE							
1	50.2	48.54	66.1	64.60	812.0	41.0	19.80	49.31	50.97	+0.77
2	50.4	48.74	65.5	64.00	735.9	43.0	17.11	49.58	51.24	+0.84
3	50.2	48.54	65.6	64.09	692.0	48.0	14.41	49.55	51.22	+1.02
4	50.0	48.34	65.0	63.48	604.8	40.2	15.04	49.28	50.94	+0.94
5	50.2	48.54	65.2	63.68	548.1	44.0	12.45	49.30	50.96	+0.76
6	50.2	48.54	66.2	64.70	500.7	41.0	12.21	49.76	51.42	+1.22
7	50.2	48.54	65.8	64.29	467.9	34.1	13.72	49.61	51.27	+1.07
8	50.5	48.84	66.9	65.42	355.8	33.0	10.78	50.25	51.90	+1.40
9	50.2	48.54	66.2	64.70	337.7	29.3	11.52	49.83	51.49	+1.29
10	50.2	48.54	66.3	64.80	301.8	27.2	11.09	49.88	51.55	+1.35
Average	50.2	48.54	65.9	64.38				49.63	51.30	+1.10
			D 21.1 MOLE PER CENT n-HEXANE IN THE PRIMARY CONDENSATE							
1	20.9	19.82	32.6	31.16	683.6	23.2	29.46	20.19	21.28	+0.38
2	21.2	20.11	32.3	30.87	594.0	36.0	16.50	20.72	21.81	+0.61
3	21.3	20.21	31.4	30.00	530.8	33.0	16.08	20.78	21.87	+0.57
4	20.7	19.63	31.8	30.38	492.5	34.0	14.48	20.32	21.41	+0.71
5	20.6	19.52	31.6	30.19	417.4	27.0	15.46	20.17	21.26	+0.66
6	20.9	19.82	32.0	30.58	344.9	24.6	14.02	20.53	21.62	+0.72
7	21.4	20.31	32.2	30.77	198.3	27.5	7.21	21.58	22.72	+1.32
8	21.6	20.50	30.3	28.92	152.3	28.0	5.44	21.81	22.95	+1.35
9	21.4	20.31	30.2	28.82	264.7	33.5	7.90	21.26	22.39	+0.99
Average	21.1	20.02	31.6	30.19				20.82	21.92	+0.81

condensate varied between 30 to 80 per cent, depending on the binary system and the composition. The fact that the enrichment of residual vapors was less than that corresponding to equilibrium with the primary condensate indicated that a vapor-phase resistance to heat and mass transfer must be present, which is true, as it will be shown later in mass transfer section. Reviewing the data in Table VII-A, B, and C, it is noticed that the degree of enrichment, defined as Murphree plate efficiency, is much higher for middle range composition than when the mixture is predominantly concentrated in one of the components.

One may also postulate that the enrichment of the residual vapors occurred through evaporation of part of the condensate in the collecting trough under the condenser, with the result that some rectification took place and the slight difference between composition of the condensate and the original vapor was due to this rectification. However, no vaporization of the condensate was observed visually.

The interfacial vapor composition was calculated by the method proposed by Kent and Pigford (discussed on pages 22 to 27), which assumes that interfacial resistance and liquid-phase resistance are absent. For this purpose Equation 49 in Chapter II was employed to calculate the interfacial vapor composition. Once the interfacial vapor composition was known, the interfacial temperature was read from the corresponding temperature-composition diagram for the appropriate binary system.

To solve Equation 49, it was required to know the height of a gas-phase transfer unit and consequently the gas-phase mass transfer coefficient. These were calculated according to the equations presented

in Chapter II and are summarized in Table V-A, B, and C for the three binary systems. The data in Table VI-A, B and C show that the magnitude of the term $(V_1/V_2)^{-H_{gl}/h}$, used in the denominator of Equation 49, was too small to be considered. Its maximum value was found to be 4.6×10^{-5} for Test No. 8 of n-hexane-toluene system. Therefore; the denominator of the right side of Equation 49, was taken as unity. This simplified Equation 49 to Equation 52, which was solved for the interfacial vapor composition, y_{AI}.

The calculated interfacial composition, along with composition of the primary condensate and vapor in equilibrium with it, as well as the composition of vapor streams entering and leaving the primary condenser are tabulated in Table VII-A, B, and C. For convenience, the average composition of various vapor streams and their differences for the three binary systems, are summarized in Table VIII.

In the derivation of Equation 49 it was assumed that there was no liquid-phase resistance to mass transfer. Had this assumption been correct, the composition of condensate from the primary condenser should have been the same as the interfacial liquid composition. Comparison of y_{AI} and y_A^* in Table VIII show that, for the three binary systems, the two composition differ from each other and that the difference between the two become greater as the concentration of more volatile component in the primary condensate decreases. Therefore, contrary to the assumption made, liquid-phase resistance to mass transfer was present and apparently its magnitude is related to the composition of the condensate. This discussion is illustrated graphically in Figures 15, 16, and 17. If there were no liquid-phase resistance present, then x_A

TABLE V

SUMMARY OF CALCULATION FOR HEIGHT EQUIVALENT TO A TRANSFER UNIT

A Benzene - n-Heptane System

Run No.	Vapor Flow Rates, Cu. Ft./Min., Condensate: Primary W	Auxiliary W_2	Total Vapor to Primary Condenser W_1	Vapor Flow Rates, Lb-Moles/Hr.: Auxiliary Condenser V_2	Primary Condenser V_1	$\frac{V_1}{V_2}$	Molal Mass Velocity of Vapor to Primary Condenser, G_{M1} Lb-Moles/Hr. Ft.^{2}	Reynolds Number of Vapor to Primary Condenser $\frac{D_e G_1'}{\mu_a}$	Mass Transfer Factor J_D	$(Sc.)^{0.56}$	Mass Transfer Coefficient For Gas Phase k_{g1} Lb-Moles/Hr. Ft.^{2} Mole Fraction	Height Equivalent to a Transfer Unit H_{G1} Ft.
				A 82.57 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	276.7	13.0	289.7	0.0211	0.4675	22.16	1.1359	780	0.0800	0.8529	0.0866	4.78
2	288.9	12.9	301.8	0.0210	0.4685	30.88	1.5795	1,080	0.0170	0.8529	0.0314	5.62
3	494.5	12.9	507.4	0.0210	0.8189	39.00	1.9896	1,370	0.0154	0.8529	0.0359	6.21
4	603.8	12.3	616.1	0.0231	0.9983	43.86	2.2458	1,660	0.0142	0.8529	0.0408	6.73
5	638.7	11.8	651.5	0.0198	1.0515	54.76	2.5546	1,760	0.0137	0.8529	0.0410	6.98
6	762.5	11.0	773.5	0.0178	1.2583	70.13	3.0330	2,090	0.0128	0.8529	0.0455	7.47
7	343.5	13.2	356.7	0.0214	0.5757	26.90	1.3986	960	0.0182	0.8529	0.0898	5.85
8	446.5	13.5	460.1	0.0219	0.7485	33.90	1.8041	1,280	0.0160	0.8529	0.0338	5.97
9	546.0	13.5	559.5	0.0219	0.9030	41.83	2.1939	1,520	0.0148	0.8529	0.0381	6.46
10	842.5	11.5	854.0	0.0187	1.3783	73.70	3.3686	2,300	0.0124	0.8529	0.0487	7.71
				B 70.37 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	248.6	12.0	260.6	0.0190	0.4078	21.43	0.9893	730	0.0808	0.8256	0.0869	4.45
2	232.7	12.6	245.3	0.0800	0.3832	19.16	0.9112	670	0.0215	0.8256	0.0868	4.30
3	306.5	13.5	318.0	0.0214	0.4969	23.22	1.2072	870	0.0190	0.8256	0.0378	4.87
4	496.4	14.9	509.3	0.0234	0.7958	33.72	1.9334	1,400	0.0152	0.8256	0.0354	6.08
5	644.6	15.2	659.8	0.0260	0.7184	29.93	1.7455	1,260	0.0160	0.8256	0.0338	5.78
6	576.8	15.3	592.1	0.0262	0.9852	38.83	2.2477	1,630	0.0143	0.8256	0.0389	6.47
7	686.8	15.8	644.6	0.0250	1.0072	40.29	2.3470	1,770	0.0137	0.8256	0.0406	6.75
8	615.6	16.0	631.6	0.0253	0.9809	39.00	2.3977	1,760	0.0138	0.8256	0.0401	6.70
9	711.0	18.0	729.0	0.0288	1.1391	39.97	2.7675	2,000	0.0112	0.8256	0.0339	7.06
10	708.3	18.0	726.3	0.0285	1.1348	39.82	2.7572	2,000	0.0111	0.8256	0.0338	7.06
11	463.2	16.0	779.2	0.0253	1.2175	48.13	2.9580	2,140	0.0127	0.8256	0.0455	7.29
12	385.5	15.1	400.6	0.0239	0.6459	26.19	1.5808	1,200	0.0170	0.8256	0.0313	5.44
				C 54.9% MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	245.0	18.0	263.0	0.0276	0.3950	14.31	0.9598	740	0.0805	0.7848	0.0869	4.19
2	336.5	17.2	353.7	0.0264	0.5313	20.18	1.2907	990	0.0178	0.7848	0.0891	4.97
3	441.3	18.0	459.3	0.0276	0.6999	25.00	1.6761	1,290	0.0158	0.7848	0.0335	5.60
4	552.5	15.5	568.0	0.0238	0.8532	35.85	2.0738	1,590	0.0145	0.7848	0.0380	6.10
5	578.3	15.5	593.8	0.0238	0.8919	37.47	2.1670	1,670	0.0140	0.7848	0.0384	6.32
6	637.0	17.0	654.0	0.0261	0.9823	37.63	2.3866	1,840	0.0135	0.7848	0.0408	6.56
7	718.5	20.0	738.5	0.0307	1.1092	36.13	2.6950	2,070	0.0128	0.7848	0.0437	6.92
8	732.9	17.5	750.4	0.0269	1.1271	41.90	2.7388	2,110	0.0127	0.7848	0.0440	6.98
9	330.2	18.1	348.3	0.0278	0.5232	18.82	1.2710	980	0.0180	0.7848	0.0289	4.98
10	580.2	16.0	596.2	0.0246	0.8955	36.40	2.1757	1,670	0.0140	0.7848	0.0385	6.32
				D 32.96 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	257.5	20.2	277.3	0.0298	0.3952	13.53	0.9603	800	0.0198	0.7647	0.0854	4.22
2	379.2	18.1	397.3	0.0266	0.5465	21.58	1.3738	1,150	0.0167	0.7647	0.0307	5.01
3	317.3	22.0	339.3	0.0318	0.4889	15.18	1.1733	980	0.0180	0.7647	0.0283	4.65
4	583.0	20.2	603.2	0.0292	0.8595	29.40	2.0858	1,750	0.0138	0.7647	0.0385	6.06
5	653.3	16.5	669.8	0.0238	0.9533	40.00	2.3161	1,940	0.0131	0.7647	0.0409	6.34
6	753.0	21.7	774.7	0.0314	1.1026	35.11	2.6709	2,240	0.0125	0.7647	0.0448	6.69
7	528.8	19.8	548.6	0.0286	0.7751	27.10	1.8832	1,580	0.0144	0.7647	0.0363	5.81
8	449.4	20.0	469.4	0.0289	0.6681	23.11	1.6233	1,360	0.0155	0.7647	0.0337	5.40
9	425.9	18.2	444.1	0.0263	0.6321	24.00	1.5357	1,290	0.0158	0.7647	0.0325	5.30
10	293.3	22.0	315.3	0.0318	0.4497	14.11	1.0903	910	0.0184	0.7647	0.0271	4.50

TABLE V

SUMMARY OF CALCULATION FOR HEIGHT EQUIVALENT TO A TRANSFER UNIT

B n-Heptane - Toluene System

Run No.	Vapor Flow Rates: Gm./Min. Condensate: Primary W	Gm./Min. Condensate: Auxiliary W_2	Gm./Min. Total Vapor to Primary Condenser W_1	Lb-Moles/Hr. Auxiliary Condenser V_2	Lb-Moles/Hr. Primary Condenser V_1	$\frac{V_1}{V_2}$	Molal Mass Velocity of Vapor to Primary Condenser, Lb-Moles/Hr. Ft.^{2} G_{M1}	Reynolds Number of Vapor to Primary Condenser $\frac{D_e G_1}{\mu_m}$	Mass Transfer Factor J_D	$(Sc.)^{0.56}$	Mass Transfer Coefficient For Gas Phase, Lb-Moles/Hr. Ft.^{2} Mole Fraction k_{g1}	Height Equivalent to a Transfer Unit, Ft. H_{G1}
				A 85.56 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	317.2	36.0	353.2	0.0680	0.4717	9.82	1.1461	1,050	0.0175	0.7528	0.0260	4.82
2	357.1	34.0	391.1	0.0654	0.5224	11.50	1.2691	1,160	0.0166	0.7528	0.0280	5.08
3	406.7	38.0	444.7	0.0907	0.5940	11.71	1.4431	1,320	0.0156	0.7528	0.0299	5.41
4	403.6	40.0	443.0	0.0934	0.5917	11.08	1.4375	1,310	0.0157	0.7528	0.0300	5.37
5	385.0	36.0	421.0	0.0680	0.5623	11.71	1.3662	1,250	0.0160	0.7528	0.0290	5.27
6	483.0	34.0	517.0	0.0654	0.6905	15.80	1.6777	1,530	0.0147	0.7528	0.0325	5.77
7	530.1	33.0	563.1	0.0680	0.7521	17.09	1.8273	1,670	0.0140	0.7528	0.0340	6.03
8	550.5	36.0	586.5	0.0680	0.7833	16.11	1.9032	1,740	0.0138	0.7528	0.0349	6.11
9	598.0	31.1	629.1	0.0635	0.8349	20.11	2.0805	1,900	0.0135	0.7528	0.0344	6.25
10	625.0	34.0	659.0	0.0654	0.8808	19.38	2.1395	1,950	0.0132	0.7528	0.0375	6.39
11	622.5	37.0	659.5	0.0698	0.8809	17.83	2.1402	1,960	0.0131	0.7528	0.0372	6.44
12	759.3	38.0	797.3	0.0907	1.0649	21.00	2.5873	2,370	0.0123	0.7528	0.0419	6.98
13	733.4	43.0	776.4	0.0974	1.0370	18.06	2.5195	2,300	0.0124	0.7528	0.0415	6.80
14	852.6	45.0	897.6	0.0600	1.1989	19.98	2.9127	2,660	0.0116	0.7528	0.0449	7.27
				B 64.13 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	421.2	36.0	457.2	0.0888	0.6815	12.73	1.5100	1,310	0.0157	0.7753	0.0306	5.53
2	523.0	32.0	555.0	0.0833	0.7548	17.42	1.8330	1,590	0.0145	0.7753	0.0343	5.99
3	521.6	38.0	559.6	0.0515	0.7607	14.77	1.8481	1,600	0.0144	0.7753	0.0343	6.01
4	468.4	38.1	506.5	0.0516	0.7414	18.24	2.2871	1,990	0.0131	0.7753	0.0386	6.63
5	727.5	38.2	765.7	0.0518	1.0610	20.10	2.5288	2,200	0.0126	0.7753	0.0411	6.90
6	725.7	38.0	765.5	0.0515	1.0606	20.80	2.5282	2,200	0.0126	0.7753	0.0411	6.90
7	861.6	41.2	902.8	0.0558	1.2272	22.00	2.9816	2,590	0.0117	0.7753	0.0450	7.48
8	569.4	39.0	608.4	0.0528	0.8270	15.66	2.0093	1,750	0.0137	0.7753	0.0355	6.34
9	990.2	41.0	1031.2	0.0595	1.4018	25.25	3.4057	2,960	0.0111	0.7753	0.0487	7.83
10	1113.0	43.0	1156.0	0.0583	1.5715	26.95	3.8180	3,320	0.0106	0.7753	0.0527	8.20
				C 50.31 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	328.1	37.6	365.7	0.0516	0.5028	9.74	1.2217	1,080	0.0176	0.7887	0.0372	5.08
2	425.8	35.5	461.3	0.0487	0.6343	13.02	1.5411	1,390	0.0158	0.7887	0.0309	5.60
3	381.0	36.6	417.6	0.0502	0.5742	11.44	1.3951	1,170	0.0165	0.7887	0.0292	5.36
4	518.0	38.2	556.2	0.0524	0.7648	14.60	1.8582	1,560	0.0145	0.7887	0.0362	6.10
5	488.5	37.0	525.5	0.0508	0.7226	14.22	1.7556	1,470	0.0150	0.7887	0.0334	5.89
6	665.7	38.0	703.7	0.0521	0.9677	18.57	2.3510	1,970	0.0131	0.7887	0.0390	6.75
7	598.8	38.0	636.8	0.0521	0.8756	16.80	2.1275	1,780	0.0137	0.7887	0.0369	6.45
8	793.8	38.0	831.8	0.0535	1.1438	21.37	2.7789	2,330	0.0124	0.7887	0.0437	7.13
9	800.0	41.0	841.0	0.0562	1.1565	20.57	2.8097	2,360	0.0120	0.7887	0.0427	7.37
10	1086.0	43.0	1127.0	0.0590	1.5497	26.26	3.7652	3,160	0.0108	0.7887	0.0515	8.18
				D 31.19 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	477.2	34.0	511.2	0.0673	0.7184	15.10	1.7357	1,180	0.0154	0.8094	0.0330	5.09
2	532.7	40.5	573.2	0.0564	0.8030	14.80	1.9648	1,550	0.0145	0.8094	0.0348	6.86
3	578.0	38.4	616.4	0.0534	0.8414	16.20	2.0989	1,670	0.0140	0.8094	0.0368	6.48
4	689.7	40.2	729.9	0.0559	1.0800	18.84	2.6788	1,970	0.0131	0.8094	0.0401	6.98
5	689.7	38.0	727.7	0.0529	1.0770	19.92	2.6708	1,970	0.0131	0.8094	0.0400	6.98
6	788.8	39.0	827.8	0.0543	1.2488	21.15	3.0903	2,280	0.0126	0.8094	0.0434	7.52
7	856.0	41.0	897.0	0.0571	1.3535	21.95	3.3456	2,480	0.0120	0.8094	0.0451	7.58
8	1150.0	43.0	1193.0	0.0598	1.8472	27.97	4.5806	3,380	0.0108	0.8094	0.0550	8.40
9	1951.4	46.0	1597.4	0.0680	2.2134	34.88	5.4837	4,380	0.0096	0.8094	0.0643	9.45
10	1480.0	39.0	1489.0	0.0543	2.0809	38.32	5.0957	4,080	0.0098	0.8094	0.0612	9.86

TABLE V

SUMMARY OF CALCULATION FOR HEIGHT EQUIVALENT TO A TRANSFER UNIT

C n-Hexane - Toluene System

Run No.	Vapor Flow Rates: Gm./Min. Condensate: Primary W	Gm./Min. Condensate: Auxiliary W_2	Gm./Min. Total Vapor to Primary Condenser W_1	lb-Moles/Hr. Auxiliary Condenser V_2	lb-Moles/Hr. Primary Condenser V_1	V_1/V_2	Molal Mass Velocity of Vapor to Primary Condenser lb-Moles/Hr. Ft.^{2} G_{M1}	Reynolds Number of Vapor to Primary Condenser $D_e G_1/\mu_m$	Mass Transfer Factor J_D	$(Sc.)^{0.56}$	Mass Transfer Coefficient for Gas Phase lb-Moles/Hr. Ft.^{2} Mole Fraction k_{g1}	Height Equivalent to a Transfer Unit Ft. H_{G1}
				A 80.40 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	848.9	43.2	892.1	0.0664	1.3530	20.39	3.8965	2,660	0.0116	0.7833	0.0408	7.43
2	845.0	39.2	884.2	0.0596	1.3484	22.57	3.3398	2,690	0.0115	0.7833	0.0490	7.40
3	688.4	38.6	727.2	0.0587	1.0982	18.80	2.6634	2,150	0.0127	0.7833	0.0432	6.88
4	636.5	29.2	665.7	0.0444	0.9779	22.08	2.3866	1,980	0.0133	0.7833	0.0405	6.57
5	563.7	37.3	601.0	0.0567	0.9108	16.05	2.2195	1,790	0.0137	0.7833	0.0388	6.39
6	533.1	32.0	565.1	0.0486	0.8528	17.54	2.0795	1,680	0.0140	0.7833	0.0369	6.89
7	501.8	37.5	539.3	0.0570	0.8167	14.32	1.9916	1,600	0.0144	0.7833	0.0364	6.07
8	360.0	25.6	375.6	0.0561	0.5688	10.51	1.3871	1,120	0.0168	0.7833	0.0297	5.20
9	299.6	11.5	311.1	0.0630	0.5166	8.20	1.2597	1,020	0.0177	0.7833	0.0285	4.94
10	388.5	36.6	425.1	0.0566	0.6438	11.37	1.5699	1,270	0.0160	0.7833	0.0321	5.46
				B 70.93 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	804.7	51.2	855.9	0.0774	1.2879	16.64	3.1456	2,490	0.0119	0.7868	0.0377	7.35
2	756.7	53.0	809.7	0.0808	1.2184	15.19	2.9758	2,350	0.0123	0.7868	0.0366	7.11
3	680.0	46.0	726.0	0.0734	1.0984	13.72	2.6682	2,110	0.0129	0.7868	0.0438	6.78
4	618.8	59.1	671.9	0.0894	1.0110	11.30	2.4694	1,950	0.0133	0.7868	0.0415	6.63
5	526.1	46.0	572.1	0.0695	0.8608	12.38	2.1086	1,660	0.0142	0.7868	0.0380	6.16
6	483.1	40.0	523.1	0.0605	0.6934	11.46	1.6946	1,340	0.0155	0.7868	0.0335	5.64
7	288.4	47.0	329.4	0.0710	0.4954	6.98	1.2106	960	0.0180	0.7868	0.0282	4.80
8	251.0	32.0	283.0	0.0488	0.4258	8.80	1.0400	820	0.0195	0.7868	0.0258	4.49
9	464.9	31.0	495.9	0.0469	0.7644	15.91	1.8825	1,460	0.0150	0.7868	0.0348	5.83
				C 51.30 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	812.0	41.0	853.0	0.0615	1.2468	20.40	3.1078	2,480	0.0121	0.7872	0.0478	7.22
2	735.9	43.0	778.9	0.0644	1.1567	17.96	2.8378	2,270	0.0126	0.7872	0.0458	6.88
3	698.0	48.0	740.0	0.0720	1.0990	15.26	2.6961	2,080	0.0129	0.7872	0.0442	6.77
4	604.8	60.2	665.0	0.0908	0.9579	15.91	2.3900	1,800	0.0137	0.7872	0.0409	6.38
5	548.1	44.0	592.1	0.0659	0.8795	13.34	2.1578	1,650	0.0143	0.7872	0.0389	6.15
6	500.7	41.0	541.7	0.0615	0.8065	13.08	2.0100	1,510	0.0148	0.7872	0.0378	5.80
7	467.9	34.1	502.0	0.0511	0.7655	14.59	1.8890	1,400	0.0152	0.7872	0.0353	5.75
8	355.8	33.0	388.8	0.0496	0.5774	11.69	1.4485	1,080	0.0172	0.7872	0.0309	5.08
9	337.7	29.3	367.0	0.0438	0.5450	11.03	1.3371	1,030	0.0177	0.7872	0.0300	4.94
10	301.8	27.2	329.0	0.0408	0.4886	11.97	1.1987	920	0.0185	0.7872	0.0287	4.72
				D 21.99 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER								
1	683.6	23.2	706.8	0.0340	1.0893	32.27	2.5170	1,860	0.0135	0.7908	0.0528	6.54
2	598.0	36.0	634.0	0.0527	0.9175	17.40	2.1434	1,660	0.0160	0.7908	0.0396	6.31
3	530.8	33.0	563.8	0.0488	0.8210	16.96	2.0077	1,490	0.0168	0.7908	0.0375	5.96
4	498.5	36.0	534.5	0.0490	0.7647	15.39	1.8768	1,390	0.0153	0.7908	0.0348	5.77
5	417.4	27.0	444.4	0.0396	0.6472	16.34	1.5825	1,170	0.0165	0.7908	0.0389	5.35
6	344.9	24.6	369.5	0.0360	0.5381	14.95	1.3158	970	0.0180	0.7908	0.0899	4.91
7	198.3	27.5	225.8	0.0403	0.3288	8.16	0.8040	600	0.0227	0.7908	0.0230	3.89
8	152.3	28.0	180.3	0.0410	0.2626	6.40	0.6480	480	0.0252	0.7908	0.0204	3.50
9	264.7	33.5	298.2	0.0491	0.4343	8.84	1.0619	790	0.0200	0.7908	0.0268	4.41

TABLE VI

SUMMARY OF CALCULATIONS FOR $(V_1 / V_2) = \frac{H_{g1}}{h}$ AND y_{AI}

A BENZENE - n-HEPTANE SYSTEM

Run No.	$\frac{V_1}{V_2}$	H_{g1} Ft.	$\frac{H_{g1}}{h}$	$(V_1 / V_2) = \frac{H_{g1}}{h}$	Mole Per Cent Benzene in Vapor y_{A1}	y_{A2}	y_{AI}
	A 82.57 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	22.16	4.78	7.35	1.29×10^{-10}	82.72	85.2	85.33
2	30.88	5.62	8.64	*	82.69	85.2	85.28
3	39.00	6.21	9.55	*	82.67	85.2	85.27
4	42.86	6.73	10.35	*	82.47	85.3	85.37
5	54.76	6.98	10.74	*	82.46	85.2	85.25
6	70.13	7.47	11.49	*	82.54	85.3	85.34
7	26.90	5.25	8.07	*	82.59	84.8	84.88
8	33.90	5.97	9.18	*	82.67	84.8	84.86
9	41.23	6.46	9.93	*	82.47	84.8	84.86
10	73.70	7.71	11.86	7.11×10^{-23}	82.44	85.2	85.24
	B 70.37 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	21.43	4.45	6.84	*	70.26	75.7	75.96
2	19.16	4.30	6.61	3.34×10^{-9}	70.50	76.0	76.30
3	23.22	4.87	7.49	*	70.21	76.0	76.26
4	33.72	6.08	9.35	*	70.37	75.8	75.96
5	29.92	5.78	8.89	*	70.39	76.2	76.60
6	38.23	6.47	9.95	*	70.35	75.8	75.95
7	40.29	6.75	10.38	*	70.35	76.2	76.35
8	39.00	6.70	10.30	*	70.46	76.4	76.55
9	39.97	7.06	10.86	*	70.34	75.7	75.84
10	39.82	7.06	10.86	*	70.44	75.7	75.83
11	48.12	7.29	11.21	1.38×10^{-19}	70.48	75.6	75.70
12	26.19	5.44	8.36	*	70.31	75.6	75.81
	C 54.94 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	14.31	4.19	6.44	3.61×10^{-8}	55.44	64.0	64.64
2	20.12	4.97	7.64	*	55.24	64.0	64.46
3	25.00	5.60	8.61	*	54.85	63.4	63.75
4	35.85	6.10	9.38	*	55.15	64.2	64.46
5	37.47	6.32	9.72	*	54.83	64.0	64.25
6	37.63	6.56	10.09	*	54.65	63.9	64.15
7	36.13	6.92	10.64	*	54.65	63.7	63.96
8	41.90	6.98	10.73	3.93×10^{-18}	54.73	64.1	64.33
9	18.82	4.92	7.56	*	54.99	63.8	64.29
10	36.60	6.32	9.72	*	54.85	64.0	64.26
	D 32.98 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	13.53	4.22	6.49	4.55×10^{-8}	33.06	38.5	38.93
2	21.58	5.01	7.70	*	32.56	38.3	38.58
3	15.18	4.65	7.15	*	33.39	39.8	40.25
4	29.40	6.06	9.32	*	33.20	40.2	40.44
5	40.00	6.34	9.75	*	33.13	40.0	40.17
6	35.11	6.69	10.29	1.25×10^{-16}	32.83	40.2	40.42
7	27.10	5.81	8.93	*	32.58	40.0	40.28
8	23.11	5.40	8.30	*	32.72	39.8	40.12
9	24.00	5.30	8.15	*	32.92	39.9	40.20
10	14.11	4.50	6.92	*	33.45	40.2	40.71

* The value of $(V_1 / V_2) = \frac{H_{g1}}{h}$ is between the two calculated figures.

TABLE VI

SUMMARY OF CALCULATIONS FOR $(V_1 / V_2) - \frac{H_{g1}}{h}$ AND y_{A1}

B n-HEPTANE-TOLUENE SYSTEM

Run No.	$\frac{V_1}{V_2}$	H_{g1} Ft.	$\frac{H_{g1}}{h}$	$(V_1 / V_2) - \frac{H_{g1}}{h}$	Mole Per Cent n-Heptane in Vapor		
					y_{A1}	y_{A2}	y_{AI}
	A 85.56 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	9.82	4.82	7.41	4.45×10^{-8}	85.70	86.7	86.81
2	11.50	5.08	7.81	*	85.70	86.6	86.68
3	11.71	5.41	8.32	*	85.62	86.7	86.79
4	11.08	5.37	8.26	*	85.53	86.7	86.80
5	11.71	5.27	8.10	*	85.78	86.7	86.77
6	15.20	5.77	8.87	*	85.70	86.7	86.76
7	17.09	6.03	9.27	*	85.66	86.7	86.76
8	16.31	6.11	9.40	*	85.66	86.7	86.76
9	20.11	6.25	9.61	*	85.64	86.6	86.65
10	19.38	6.39	9.83	*	85.70	86.6	86.65
11	17.83	6.44	9.90	*	85.75	86.6	86.65
12	21.00	6.92	10.64	*	85.72	86.6	86.64
13	18.06	6.80	10.46	*	85.66	86.7	86.76
14	19.98	7.27	11.18	2.88×10^{-15}	85.66	86.7	86.75
	B 64.13 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	12.73	5.53	8.50	4.06×10^{-10}	64.40	67.7	67.98
2	12.42	5.99	9.21	*	64.32	67.8	68.01
3	14.77	6.03	9.27	*	64.34	67.7	67.94
4	18.24	6.63	10.20	*	64.12	67.8	68.01
5	20.10	6.90	10.61	*	64.10	67.7	67.89
6	20.20	6.90	10.61	*	64.13	67.7	67.88
7	22.00	7.42	11.41	*	63.89	67.8	67.98
8	15.66	6.34	9.75	*	64.06	67.8	68.05
9	25.25	7.83	12.04	*	63.97	67.8	67.96
10	26.95	8.20	12.61	9.13×10^{-19}	63.96	67.8	67.95
	C 50.32 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	9.74	5.02	7.72	2.33×10^{-8}	50.63	53.2	53.49
2	13.02	5.60	8.61	*	50.26	53.2	53.44
3	11.44	5.36	8.24	*	50.28	53.2	53.48
4	14.60	6.10	9.38	*	50.58	53.1	53.28
5	14.22	5.89	9.06	*	50.58	53.1	53.29
6	18.57	6.75	10.38	*	50.15	52.8	52.95
7	16.80	6.45	9.92	*	50.17	52.9	53.07
8	21.37	7.13	10.97	*	50.14	53.0	53.14
9	20.57	7.37	11.33	*	50.43	53.1	53.23
10	26.26	8.18	12.58	1.39×10^{-18}	50.02	53.2	53.32
	D 31.19 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	15.10	5.89	9.06	2.08×10^{-11}	31.53	35.9	36.21
2	14.20	6.26	9.63	*	31.52	35.8	36.12
3	16.10	6.48	9.97	*	31.33	35.7	35.99
4	18.24	6.92	10.64	*	31.08	36.0	36.28
5	19.22	6.92	10.64	*	31.43	35.8	36.03
6	21.15	7.20	11.07	*	31.53	35.8	36.01
7	21.95	7.56	11.63	*	30.26	35.8	36.06
8	27.87	8.40	12.92	*	31.55	35.8	35.96
9	34.88	9.45	14.53	*	30.26	35.8	35.95
10	38.32	9.26	14.24	2.83×10^{-23}	31.02	35.8	35.93

* The value of $(V_1 / V_2) - \frac{H_{g1}}{h}$ is between the two calculated figures.

TABLE VI

SUMMARY OF CALCULATIONS FOR $(V_1 / V_2) - \frac{H_{g1}}{h}$ AND y_{AI}

C n-HEXANE - TOLUENE SYSTEM

Run No.	$\frac{V_1}{V_2}$	H_{P_1} Ft.	$\frac{\mu_{P_1}}{h}$	$(V_1 / V_2) - \frac{H_{g1}}{h}$	Mole Per Cent n-Hexane in Vapor y_{A1}	y_{A2}	y_{AI}
	A 80.10 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	20.59	7.63	11.74	3.78×10^{-16}	80.23	86.6	86.92
2	22.07	7.60	11.70	*	80.64	86.0	86.24
3	18.60	6.88	10.58	*	80.15	86.2	86.54
4	22.02	6.57	10.10	*	79.96	85.7	85.97
5	16.05	6.39	9.83	*	80.61	86.2	86.50
6	17.54	6.29	9.67	*	80.56	81.9	85.16
7	14.32	6.07	9.33	*	80.58	81.8	85.11
8	10.51	5.30	8.00	*	80.27	86.2	86.82
9	8.20	4.94	7.60	1.13×10^{-7}	80.10	86.1	86.89
10	11.37	5.16	8.10	*	80.62	86.0	86.51
	B 70.93 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	16.64	7.35	11.30	1.59×10^{-14}	70.68	78.3	78.79
2	15.19	7.11	10.94	*	70.75	78.6	79.15
3	13.72	6.78	10.43	*	70.58	78.0	78.58
4	11.30	6.63	10.20	*	70.71	79.3	80.13
5	12.38	6.16	9.47	*	71.23	78.0	79.68
6	11.46	5.64	8.67	*	71.07	78.7	79.43
7	6.98	4.80	7.38	5.93×10^{-7}	71.55	78.3	79.43
8	8.80	4.49	6.90	*	71.94	78.6	79.57
9	15.91	5.83	8.97	*	70.77	79.5	80.08
	C 51.30 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	20.60	7.22	11.10	2.60×10^{-15}	50.97	66.1	66.87
2	17.96	6.93	10.66	*	51.24	65.5	66.34
3	15.26	6.77	10.41	*	51.22	65.6	66.60
4	15.91	6.38	9.81	*	50.94	65.0	65.94
5	13.34	6.15	9.46	*	50.96	65.2	66.35
6	13.08	5.80	8.92	*	51.42	66.2	67.42
7	14.59	5.75	8.84	*	51.27	65.8	66.87
8	11.69	5.08	7.81	*	51.90	66.9	68.30
9	11.03	4.93	7.58	*	51.49	66.2	67.66
10	11.97	4.72	7.26	1.49×10^{-8}	51.55	66.3	67.64
	D 21.92 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER						
1	30.27	6.54	10.06	1.22×10^{-15}	21.28	32.6	32.98
2	17.40	6.31	9.70	*	21.81	32.3	32.94
3	16.96	5.96	9.17	*	21.87	31.4	32.00
4	15.39	5.77	8.87	*	21.41	31.8	32.52
5	16.34	5.35	8.23	*	21.26	31.6	32.27
6	14.95	4.91	7.55	*	21.62	32.0	32.74
7	8.16	3.89	5.98	*	22.72	32.2	33.52
8	6.40	3.50	5.38	4.60×10^{-5}	22.95	30.3	31.66
9	8.84	4.41	6.78	*	22.39	30.2	31.20

* The value of $(V_1 / V_2) - \frac{H_{g1}}{h}$ is between the two calculated figures.

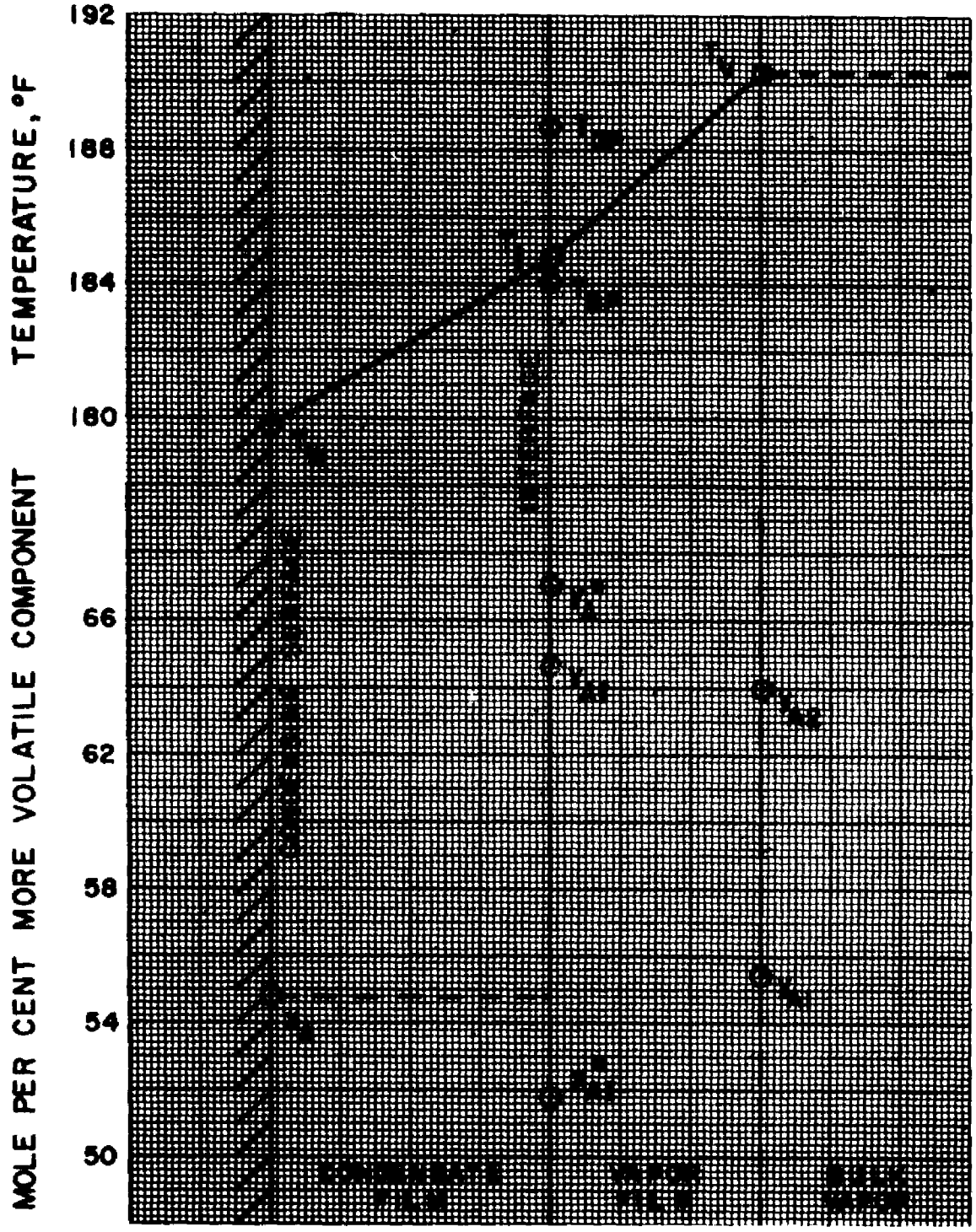


FIGURE 15. TEMPERATURE AND COMPOSITION DROP THROUGH THE FILMS

DATA FOR TEST NO. 1, MIXTURE C, OF BENZENE—n-HEPTANE SYSTEM

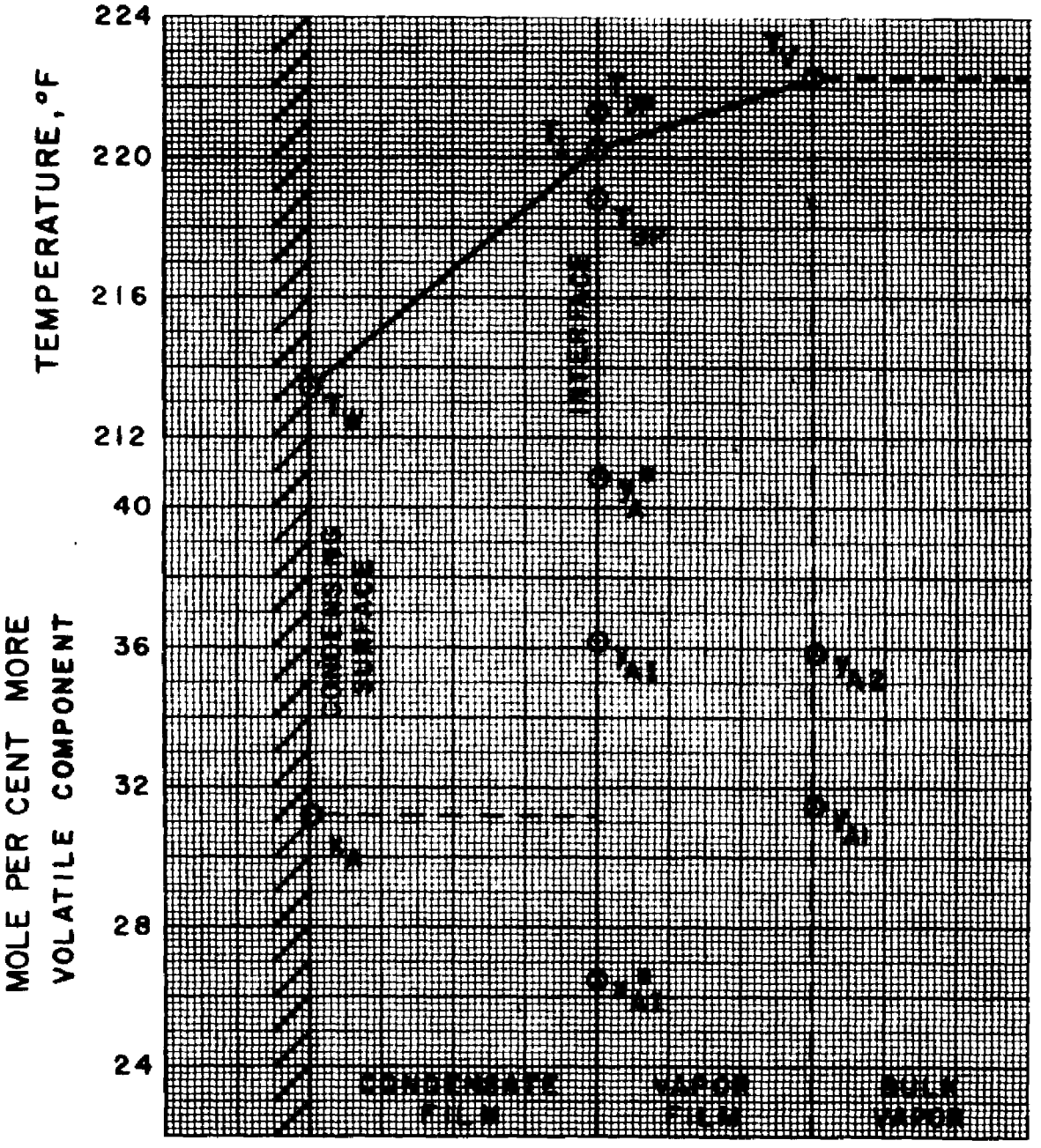


FIGURE 16. TEMPERATURE AND COMPOSITION DROP THROUGH THE FILMS

DATA FOR TEST NO. 1, MIXTURE D OF n-HEPTANE-TOLUENE SYSTEM

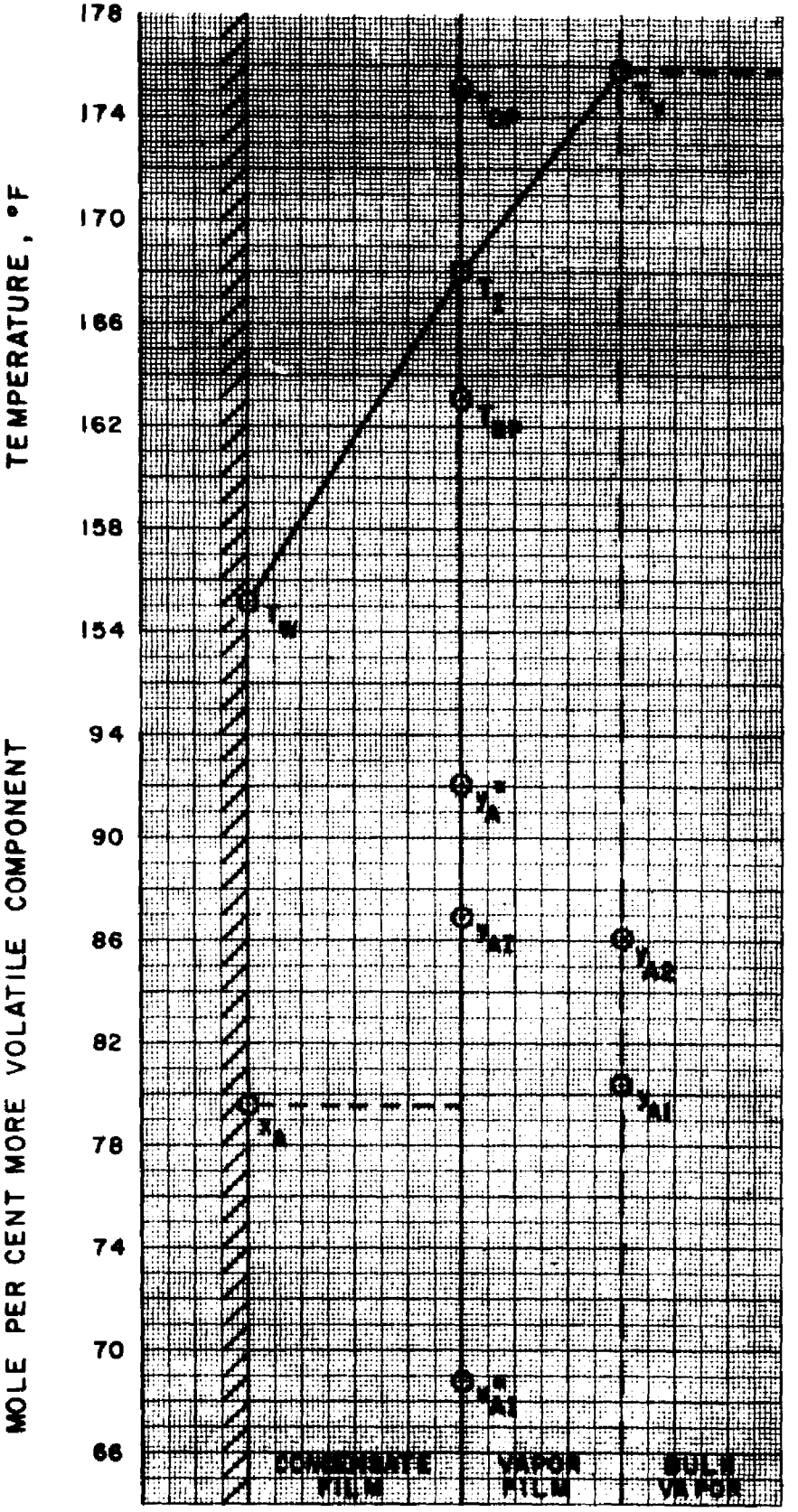


FIGURE 17. TEMPERATURE AND COMPOSITION DROPS THROUGH THE FILMS

DATA FOR TEST NO. 9, MIXTURE A OF n-HEXANE-TOLUENE SYSTEM

should have coincided with x^*_{AI} and y_{A^*} should have been the same as y_{AI}. The fact that these points differ from each other indicate the existence of liquid-phase resistance. Also if there were no gas-phase resistance present, then y_{AI} should have been the same as y_{Al}. Again the difference between the two is indication of presence of gas-phase resistance.

The interfacial temperature, vapor temperatue, dew point and bubble point temperatures of the condensate for mixtures of the three binary systems are tabulated in Table VII-A, B and C. The average temperatures for each mixture are summarized in Table IX. It is noticed from the latter table that for all cases, except one, the interface temperature is closer to the boiling point of the condensate than to either the dew point of the condensate or the vapor temperature. The temperature difference between the vapor and the interfacial temperature again indicates the existence of the gas-phase resistance. Otherwise the interfacial temperature would have been the same as the vapor temperature. The one exception indicated above, in which the interfacial temperature is closer to the dew point of the mixture than to its boiling point, is the n-hexane-toluene mixture with a condensate composition of 21.1 mole per cent n-hexane. The mass transfer calculation for this particular mixture, which will be discussed in a later paragraph, indicates that its gas-phase resistance is much lowre than those for the other mixtures of n-hexane-toluene system. This factor explains why the interfacial temperature is closer to the dew point than to the bubble point of the condensate.

It is also noticed from Table IX that the difference between the

TABLE VII

COMPARISON OF HEAT TRANSFER COEFFICIENTS BASED ON INTERFACIAL TEMPERATURE WITH THOSE BASED ON VAPOR, DEW POINT, AND BOILING POINT TEMPERATURES

A BENZENE - n-HEPTANE SYSTEM

Run No.	Mole Per Cent Benzene in Mixture						Murphree Efficiency Per Cent	Temperature, °F.					Temperature Drop, °F.				Heat Transfer Coefficient Btu/hr. ft.² °F.			
	x_A	y^*_A	y_{A1}	y_{A2}	y_{AI}	x^*_{A1}	E^*_{MV}	t_M	t_V	t_{DP}	t_{BP}	t_I	Δt_V	Δt_{DP}	Δt_{BP}	Δt_I	h_V	h_{DP}	h_{BP}	h_I
A 82.57 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER																				
1	82.6	86.6	82.72	85.2	85.31	81.3	63.9	181.6	179.8	179.0	179.2	178.4	11.7	10.9	10.1	10.3	307	331	359	347
2	82.6	86.6	82.66	85.2	85.28	81.3	64.2	166.1	179.8	179.0	179.2	178.4	11.7	10.9	10.1	10.3	307	331	359	347
3	82.6	86.6	82.67	85.2	85.27	81.3	64.4	163.2	179.9	179.0	179.2	178.4	16.7	15.8	15.0	15.2	273	289	305	300
4	82.4	86.4	82.47	85.3	85.37	81.3	72.0	157.1	179.9	179.0	179.2	178.3	22.8	21.9	21.1	21.2	249	254	263	261
5	82.4	86.4	82.46	85.2	85.25	81.3	69.5	164.3	179.9	179.0	179.2	178.4	26.6	25.7	23.9	24.1	230	239	247	245
6	82.5	86.4	82.50	85.1	85.34	81.2	71.5	157.6	179.8	179.0	179.2	178.4	32.2	31.4	30.6	30.8	212	216	230	228
7	82.5	86.4	82.59	86.4	84.88	81.2	58.0	177.8	179.9	179.0	178.2	178.4	9.1	8.7	7.4	7.6	268	307	328	313
8	82.6	86.6	82.67	86.8	84.86	81.2	58.1	166.6	179.8	179.1	178.2	178.4	14.7	13.6	12.6	12.8	290	306	328	321
9	82.4	86.4	82.47	86.8	84.86	81.2	59.2	159.9	179.9	179.0	178.2	178.4	20.0	19.0	18.1	18.5	252	265	277	272
10	82.4	86.4	82.44	85.2	85.24	81.3	59.7	161.3	179.8	179.1	178.2	178.4	38.5	37.8	36.9	37.1	202	206	211	209
Average	82.5	86.5	82.57	85.1	85.17	81.2	64.6		179.8	179.0	178.2	178.4								
B 70.37 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER																				
1	70.0	77.6	70.26	75.7	75.96	67.5	74.1	175.9	184.8	183.0	180.4	180.9	9.9	7.1	4.5	5.0	253	317	500	444
2	70.2	77.7	70.50	76.0	76.30	68.0	76.7	175.9	184.0	183.0	180.4	180.9	8.1	7.1	4.5	5.0	262	297	449	444
3	70.0	77.6	70.21	76.2	76.26	68.0	81.0	172.8	184.3	183.0	180.4	180.9	11.5	10.2	7.6	8.1	260	271	363	363
4	70.2	77.7	70.37	75.8	75.96	67.5	74.0	166.2	184.6	182.9	180.4	180.9	19.4	17.7	15.2	15.7	210	230	269	257
5	70.2	77.7	70.39	76.2	76.40	68.1	79.5	166.0	184.1	183.0	180.3	180.9	19.1	18.0	15.4	15.8	211	228	268	256
6	70.3	77.7	70.35	75.8	75.95	67.5	73.9	157.6	184.8	182.9	180.4	180.9	27.2	25.3	22.8	23.3	192	206	230	223
7	70.2	77.7	70.35	76.2	76.35	68.0	79.6	154.7	184.2	182.9	180.3	180.9	29.9	28.6	26.0	26.5	190	198	218	213
8	70.3	77.8	70.46	76.4	76.55	68.4	80.9	154.2	182.2	180.9	180.3	180.7	30.0	28.6	26.1	26.5	186	195	214	209
9	70.2	77.7	70.34	75.7	75.84	67.6	72.8	149.3	184.1	182.9	180.3	180.9	34.8	33.6	31.0	31.6	185	192	208	202
10	70.3	77.8	70.44	75.7	75.83	67.6	71.4	149.5	184.7	182.9	180.3	180.9	35.2	33.4	30.8	31.4	182	192	208	203
11	70.3	77.8	70.48	75.6	75.70	67.2	69.9	148.1	184.1	182.9	180.3	181.0	20.0	18.8	16.2	16.9	210	223	259	247
12	70.1	77.7	70.31	75.6	75.80	67.2	71.6	159.2	184.7	183.0	180.3	181.0	15.5	13.8	11.1	11.8	225	253	332	295
Average	70.2	77.7	70.37	75.9	76.08	67.7		164.4	184.4	182.9	180.3	180.9								
C 59.9% MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER																				
1	54.8	67.0	55.44	64.0	64.64	51.8	74.2	179.8	190.3	188.7	184.0	184.8	10.5	8.9	4.2	5.0	206	244	514	430
2	54.8	67.0	55.24	64.0	64.46	51.5	74.5	175.5	190.4	188.7	184.0	184.9	14.9	13.2	8.5	9.4	200	226	395	334
3	54.5	66.8	54.85	63.4	63.75	50.7	71.5	169.5	190.6	188.8	184.1	185.1	21.1	19.3	14.6	15.6	185	202	288	268
4	54.9	67.0	55.15	64.2	64.46	51.5	76.3	162.6	191.2	188.7	184.0	184.9	28.6	26.0	21.3	22.3	171	188	230	218
5	54.6	66.8	54.83	64.0	64.25	51.3	76.6	162.0	190.6	188.8	184.1	184.9	28.6	26.8	22.1	22.9	179	191	232	221
6	54.4	66.6	54.65	63.9	64.15	51.2	77.4	158.2	191.0	188.8	184.1	185.0	32.8	30.6	25.9	26.8	172	184	218	208
7	54.4	66.6	54.66	63.7	63.96	51.0	75.7	153.4	190.7	188.8	184.1	185.0	37.3	35.4	30.7	31.6	171	180	207	200
8	54.5	66.7	54.73	64.1	64.33	51.2	78.3	154.2	192.1	188.8	184.1	185.0	37.9	36.6	29.9	30.8	171	187	210	208
9	54.5	66.7	54.99	63.8	64.30	51.2	75.2	175.1	190.3	188.8	184.1	185.0	15.1	13.5	8.8	9.7	194	216	332	300
10	54.6	66.8	54.85	64.0	64.25	51.2	76.5	161.6	190.7	188.8	184.1	185.0	29.1	27.2	22.5	23.4	176	189	228	218
Average	54.6	66.8	54.94	63.9	64.25	51.2	75.6	160.8	190.8	188.8	184.1	185.0								
D 32.9% MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER																				
1	32.6	48.6	33.06	38.5	38.93	24.0	35.0	186.7	198.8	198.0	191.1	195.3	12.1	11.3	4.4	8.6	180	194	493	247
2	32.3	48.2	32.56	38.3	38.58	23.7	34.7	180.7	198.9	198.1	191.2	195.4	18.2	17.4	10.5	14.7	176	184	368	211
3	32.8	48.8	33.39	39.8	40.25	25.0	41.6	183.2	198.5	197.8	191.0	194.8	15.3	14.7	7.8	11.6	175	183	343	226
4	32.8	48.8	33.20	40.2	40.44	25.2	44.9	170.3	198.6	197.8	191.2	194.6	28.3	27.5	20.9	24.3	174	179	238	198
5	32.8	48.8	33.13	40.0	40.17	25.0	43.8	166.3	198.5	197.8	191.0	194.7	32.2	31.5	24.7	28.4	171	175	223	190
6	32.6	48.7	32.83	40.2	40.42	25.2	44.4	160.2	198.5	197.9	191.1	194.6	38.3	37.7	30.9	34.4	166	168	206	181
7	32.7	48.2	32.98	40.0	40.28	25.1	47.5	154.2	198.7	198.1	191.2	194.7	34.5	33.9	27.0	30.5	181	185	249	213
8	32.5	48.5	32.72	39.8	40.12	25.0	44.8	177.2	198.6	198.0	191.2	194.7	21.4	20.8	14.0	17.5	177	182	370	211
9	32.6	48.6	32.92	39.9	40.20	25.2	44.5	179.1	198.7	197.9	191.1	194.7	19.6	18.8	12.0	15.6	183	190	398	235
10	32.8	48.8	32.65	40.2	40.71	25.5	47.4	185.4	198.9	197.9	191.0	194.5	13.5	12.5	5.6	9.2	183	198	462	266
Average	32.6	48.6	32.88	39.7	40.00	24.9	43.3		198.7	198.0	191.2	194.8								

TABLE VII

COMPARISON OF HEAT TRANSFER COEFFICIENTS BASED ON INTERFACIAL TEMPERATURES WITH THOSE BASED ON VAPOR, DEW POINT, AND BOILING POINT TEMPERATURES

3 n-HEPTANE - TOLUENE SYSTEM

Run No.	Mole Per Cent Heptane in Mixture						Murphree Efficiency Per Cent	Temperature, °F.					Temperature Drop, °F.				Heat Transfer Coefficient Btu/hr. ft.² °F.			
	x_1	y^*_1	y_{11}	y_{12}	y_{1T}	y^*_{12}	E_{MV}	t_V	t_I	t_{DP}	t_{BP}	t_T	Δt_V	Δt_{DP}	Δt_{BP}	Δt_T	h_V	h_{DP}	h_{BP}	h_T
	A 85.48 MOLE PER CENT n-HEPTANE IN y_1, ENTERING VAPOR TO THE PRIMARY CONDENSER																			
1	85.6	88.2	85.70	86.7	86.81	83.8	60.0	202.7	210.8	210.4	210.3	210.3	8.1	7.7	7.6	7.6				
2	85.6	88.2	85.70	86.6	86.68	83.7	56.0	202.0	210.9	210.4	210.3	210.3	9.9	9.4	9.3	9.3				
3	85.5	88.2	85.62	86.7	86.79	83.8	59.6	197.7	210.8	210.4	210.3	210.3	13.1	12.7	12.6	12.6				
4	85.4	88.1	85.53	86.7	86.80	83.8	53.3	196.6	210.8	210.4	210.3	210.3	14.2	13.8	13.7	13.7				
5	85.7	88.3	85.78	86.7	86.77	83.8	64.4	198.2	210.7	210.4	210.2	210.3	12.5	12.2	12.1	12.1				
6	85.6	88.2	85.70	86.7	86.76	83.8	58.0	189.5	210.9	210.4	210.3	210.3	21.4	20.9	20.8	20.8				
7	85.6	88.2	85.66	86.7	86.76	83.8	60.9	184.9	210.9	210.4	210.3	210.3	26.0	25.5	25.4	25.4				
8	85.6	88.2	85.66	86.7	86.76	83.7	60.9	185.1	211.2	210.4	210.3	210.3	26.1	25.3	25.2	25.2				
9	85.6	88.2	85.68	86.6	86.68	83.7	57.5	180.8	211.0	210.4	210.3	210.3	31.0	30.4	30.3	30.3				
10	85.7	88.3	85.70	86.6	86.68	83.7	56.6	176.8	210.9	210.4	210.3	210.3	34.1	33.4	33.3	33.3				
11	85.7	88.3	85.75	86.6	86.69	83.7	53.3	177.0	210.8	210.4	210.3	210.3	33.8	33.4	33.3	33.3				
12	85.7	88.3	85.72	86.6	86.68	83.7	54.1	172.9	210.8	210.4	210.3	210.3	37.9	37.5	37.4	37.4				
13	85.6	88.2	85.66	86.7	86.76	83.8	60.9	173.5	210.9	210.4	210.3	210.3	37.4	36.9	36.8	36.8				
14	85.6	88.2	85.66	86.7	86.75	83.8	60.9	170.5	210.8	210.4	210.3	210.3	40.3	39.9	39.8	39.8				
Average	85.6	88.2	85.68	86.7	86.72	83.8	60.5		210.9	210.4	210.3	210.3								
	B 64.13 MOLE PER CENT n-HEPTANE IN y_1, ENTERING VAPOR TO THE PRIMARY CONDENSER																			
1	64.1	69.2	64.10	67.7	67.96	62.5	68.8	203.9	213.8	213.5	212.5	212.8	9.9	9.6	8.6	8.9				
2	64.1	69.2	64.32	67.8	68.01	62.5	71.3	199.3	213.8	213.5	212.5	212.8	14.5	14.2	13.2	13.5				
3	64.1	69.2	64.34	67.7	67.96	62.5	69.1	199.2	213.8	213.5	212.5	212.8	14.6	14.3	13.3	13.6				
4	63.9	69.1	64.12	67.8	68.01	62.5	73.9	192.4	213.8	213.5	212.5	212.8	21.4	21.1	20.1	20.4				
5	63.9	69.1	64.10	67.7	67.99	62.4	72.0	187.2	213.8	213.5	212.5	212.8	26.6	26.3	25.3	25.6				
6	63.9	69.1	64.13	67.7	67.98	62.4	71.8	187.0	213.8	213.5	212.5	212.8	26.8	26.5	25.5	25.6				
7	63.7	69.1	63.89	67.8	67.98	62.5	75.0	188.4	213.8	213.6	212.5	212.8	25.4	25.2	24.1	24.4				
8	63.8	69.0	64.04	67.8	68.05	62.5	75.7	196.8	213.8	213.5	212.5	212.8	17.0	16.7	15.7	16.0				
9	63.8	69.0	63.97	67.8	67.96	62.5	76.1	179.2	213.8	213.6	212.5	212.8	34.6	34.4	33.3	33.6				
10	63.8	69.0	63.96	67.8	67.95	62.5	76.2	175.5	213.8	213.6	212.5	212.8	38.3	38.1	37.0	37.3				
Average	63.9	69.1	64.13	67.8	67.96	62.5	73.8		213.8	213.6	212.5	212.8								
	C 50.32 MOLE PER CENT n-HEPTANE IN y_1, ENTERING VAPOR TO THE PRIMARY CONDENSER																			
1	50.4	58.0	50.65	53.2	53.49	45.0	34.9	209.7	216.8	216.5	216.8	215.8	7.1	6.8	5.1	6.1				
2	50.0	57.7	50.26	53.2	53.44	44.9	39.5	205.2	217.0	216.6	216.9	215.9	11.8	11.4	9.7	10.7				
3	50.0	57.7	50.28	53.2	53.46	45.0	39.3	207.6	217.1	216.6	216.9	215.8	9.5	9.0	7.3	8.2				
4	50.4	58.0	50.58	53.1	53.38	44.5	34.8	200.8	216.8	216.5	216.8	215.9	14.8	14.5	14.8	15.9				
5	50.4	58.0	50.58	53.1	53.39	44.5	32.6	201.4	216.8	216.5	216.8	215.9	15.4	15.1	13.4	14.5				
6	50.0	57.7	50.25	52.8	52.95	44.2	35.1	193.7	217.0	216.6	216.9	216.0	23.3	22.8	21.2	22.3				
7	50.0	57.7	50.17	52.9	53.07	44.6	36.0	195.4	217.0	216.6	216.9	215.9	21.6	21.2	19.5	20.5				
8	50.0	57.7	50.14	53.0	53.14	44.7	37.8	191.2	217.0	216.6	216.9	215.9	25.8	25.4	23.7	24.7				
9	50.3	57.9	50.43	53.1	53.23	44.8	35.7	190.6	216.9	216.5	216.8	215.9	26.3	25.9	24.2	25.3				
10	49.9	57.7	50.08	53.2	53.32	44.9	41.4	187.0	217.0	216.6	216.9	215.9	30.0	29.6	27.9	28.9				
Average	50.1	57.8	50.32	53.0	53.27	44.7	35.8		216.9	216.6	216.9	215.9								
	D 31.19 MOLE PER CENT n-HEPTANE IN y_1, ENTERING VAPOR TO THE PRIMARY CONDENSER																			
1	31.2	38.9	31.53	35.9	36.21	36.5	66.6	213.4	222.3	221.3	218.8	220.2	8.9	7.9	5.4	6.8				
2	31.2	38.9	31.52	35.8	36.12	36.5	65.6	211.4	222.4	221.4	218.9	220.2	11.0	10.0	7.5	8.6				
3	31.0	38.7	31.33	35.7	35.99	36.3	64.6	209.6	222.3	221.4	218.9	220.3	12.7	12.0	9.3	10.7				
4	30.8	38.6	31.08	36.0	36.38	36.5	51.7	204.9	222.8	221.6	219.0	220.2	17.9	16.7	14.1	15.3				
5	31.2	38.9	31.43	35.8	36.03	36.3	64.1	204.7	222.5	221.7	219.0	220.3	17.8	17.0	14.3	15.6				
6	31.3	38.9	31.53	35.8	36.08	36.3	65.6	198.9	222.1	221.6	218.9	220.3	23.2	22.7	20.0	21.4				
7	30.9	38.6	31.26	35.8	36.06	36.3	56.9	195.0	222.6	221.5	218.8	220.3	27.6	26.5	23.8	25.3				
8	31.4	39.0	31.55	35.8	35.96	36.3	64.9	191.5	222.6	221.9	219.2	220.3	31.1	30.4	27.7	28.8				
9	30.5	38.0	30.88	35.8	35.95	36.3	55.0	186.9	222.4	221.8	219.1	220.3	35.5	35.6	32.9	34.1				
10	30.9	38.4	31.08	35.8	35.93	36.3	59.9	187.2	222.2	221.8	219.0	220.3	35.0	34.0	31.8	33.1				
Average	31.0	38.6	31.19	35.8	36.05	36.3	59.0		222.5	221.6	218.9	220.3								

TABLE VII

COMPARISON OF HEAT TRANSFER COEFFICIENTS BASED ON INTERFACIAL TEMPERATURE WITH THOSE BASED ON VAPOR, DEW POINT, AND BOILING POINT TEMPERATURES

C n-HEXANE - TOLUENE SYSTEM

Run No.	Mole Per Cent Hexane in Mixture					Murphree Efficiency Per Cent	Temperature, °F.					Temperature Drop, °F.				Heat Transfer Coefficient Btu/hr. Ft.^{2} °F.			
	x_A	y^*_A	y_{A1}	y_{A2}	y_{A1}	x^*_{A1}	E'_{MV}	T_V	T_V	T_{DP}	T_{BP}	T_I	Δt_V	Δt_{DP}	Δt_{BP}	Δt_I	h_V	h_{DP}	h_{BP}	h_I
					A	80.4C MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER													
1	79.9	92.2	80.22	86.6	86.92	68.9	53.2	130.4	176.7	175.0	163.0	160.0	46.3	44.6	32.6	37.6	159	165	225	199
2	80.4	92.4	80.64	86.0	86.24	67.8	45.6	128.5	175.8	174.4	162.8	162.5	47.3	45.9	34.3	40.0	158	163	218	190
3	79.8	92.2	80.15	86.2	86.54	68.0	50.2	137.9	176.1	175.0	163.0	160.4	38.2	37.1	25.1	30.5	156	159	225	197
4	79.7	92.0	79.96	85.7	85.97	67.0	47.7	141.6	176.6	175.2	163.1	160.0	35.0	33.6	21.5	27.9	152	159	245	198
5	80.3	92.4	80.61	85.2	85.50	66.0	38.9	143.3	175.8	174.6	162.8	160.3	32.5	31.3	19.5	26.0	150	156	238	191
6	80.3	92.4	80.56	84.9	85.16	65.6	36.6	144.6	175.2	174.6	162.8	160.2	30.6	30.0	18.2	24.6	150	153	232	190
7	80.3	92.4	80.58	84.8	85.11	65.6	35.7	146.3	176.0	174.6	162.8	160.2	29.7	28.3	16.5	22.9	146	153	243	193
8	79.9	92.2	80.27	86.2	86.82	68.7	49.7	153.2	175.4	175.0	163.0	160.0	22.2	21.8	9.8	14.8	132	136	188	202
9	79.6	92.0	80.40	86.1	86.89	68.8	49.1	155.2	175.8	175.1	163.1	160.0	20.6	19.9	7.9	12.8	126	130	188	206
10	80.1	92.3	80.62	86.0	86.51	68.1	47.3	150.9	175.3	174.6	162.9	160.3	24.4	23.7	12.0	17.4	138	142	200	197
Average	80.0	92.2	80.40	85.8	86.16	67.5	45.7		175.9	174.8	162.9	160.6								
					B	70.93 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER													
1	70.2	87.6	70.68	78.3	79.29	54.2	45.0	140.2	185.2	183.5	167.4	175.7	45.0	43.3	27.2	35.0	157	169	268	206
2	70.2	87.6	70.75	78.6	79.15	54.1	46.6	142.8	184.9	183.5	167.4	175.2	42.1	40.7	24.6	32.4	158	169	278	210
3	70.1	87.5	70.58	78.0	78.58	52.7	43.8	146.2	184.8	183.7	167.5	176.0	38.6	37.5	21.3	29.8	155	159	289	205
4	69.9	87.5	70.71	79.3	80.13	56.0	51.1	148.0	185.6	183.9	167.6	176.3	36.6	34.9	18.6	25.3	147	154	289	217
5	70.6	87.7	71.22	79.0	79.68	54.7	47.2	151.9	184.7	183.2	167.3	175.0	32.8	31.3	15.4	23.1	141	146	288	208
6	70.4	87.6	71.07	78.7	79.43	54.2	40.1	156.1	185.0	183.4	167.3	175.2	28.9	27.3	11.2	19.1	128	135	288	208
7	70.5	87.7	71.55	78.3	79.43	52.2	38.7	161.3	185.3	183.3	167.3	175.2	24.0	22.0	6.2	13.9	103	113	288	188
8	70.1	87.5	71.04	78.6	79.57	54.7	39.8	162.4	184.7	183.7	167.5	175.0	22.3	21.3	5.1	12.6	99	103	288	179
9	70.2	87.6	70.77	79.5	80.08	55.5	51.9	154.2	185.8	183.5	167.4	176.5	31.6	29.3	13.2	20.3	129	139	288	206
Average	70.2	87.6	70.93	78.7	79.48	54.5	46.6		185.1	183.5	167.4	175.1								
					C	51.30 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER													
1	50.2	77.0	50.97	66.1	66.87	35.7	58.1	156.2	201.2	200.1	177.4	186.6	45.0	43.9	21.2	30.4	163	167	346	285
2	50.4	77.2	51.24	65.5	66.34	35.0	54.9	158.3	201.3	200.0	177.5	187.0	43.0	41.7	19.0	28.7	155	159	350	295
3	50.2	77.0	51.22	65.6	66.40	34.3	55.8	160.0	201.3	199.5	176.9	186.8	41.3	39.5	16.9	26.8	151	158	370	291
4	50.0	76.8	50.94	65.0	65.84	34.5	56.1	163.0	201.7	200.0	177.3	187.5	38.7	37.0	14.3	26.5	141	144	384	284
5	50.2	77.0	50.96	65.2	66.35	35.0	58.7	165.9	201.9	200.1	177.4	187.0	35.0	34.1	11.5	22.9	141	145	431	329
6	50.2	77.0	51.42	66.2	67.42	36.1	57.7	167.1	201.4	200.1	177.4	186.7	34.3	33.0	10.3	19.1	132	137	430	326
7	50.2	77.0	51.27	65.8	66.87	35.7	56.5	168.4	200.8	200.1	177.4	186.6	32.4	31.7	9.0	18.2	130	133	430	316
8	50.5	77.4	51.90	66.9	68.30	37.5	58.8	171.6	201.6	199.9	177.2	185.4	30.0	28.3	5.6	13.8	107	116	358	256
9	50.2	77.0	51.49	66.2	67.66	37.3	57.6	171.7	201.7	200.1	177.4	185.8	30.0	28.4	5.7	14.1	108	107	358	280
10	50.8	77.0	51.55	66.3	67.84	37.3	57.9	172.7	200.2	199.5	176.9	185.8	27.5	26.8	4.2	13.1	99	102	400	211
Average	50.2	77.0	51.30	65.9	67.00	35.8	56.8		201.2	200.0	177.3	186.5								
					D	21.92 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER													
1	20.9	51.0	21.26	32.6	32.98	11.0	38.1	173.4	221.4	219.6	199.8	212.3	48.0	46.2	26.4	38.9	132	137	280	181
2	21.2	51.3	21.81	32.3	32.94	11.0	35.6	177.6	221.7	219.5	199.6	212.3	44.1	41.9	22.0	34.7	125	131	259	157
3	21.3	51.3	21.87	31.4	32.00	10.5	32.4	180.6	220.6	219.5	199.5	212.9	40.0	38.9	18.9	32.3	123	126	289	150
4	20.7	50.7	21.41	31.8	32.52	10.7	35.5	182.8	220.9	219.7	200.0	212.6	38.1	36.9	17.2	29.8	120	124	288	151
5	20.6	50.5	21.26	31.6	32.27	10.7	35.3	186.2	221.3	219.8	200.0	212.7	35.1	33.6	13.8	26.5	116	115	288	144
6	20.9	51.0	21.62	32.0	32.76	11.0	35.3	189.5	221.1	219.6	199.8	212.3	31.6	30.1	10.3	22.8	101	106	288	136
7	21.4	51.5	22.72	32.2	33.52	10.7	32.9	194.2	221.8	219.4	199.3	212.6	27.6	25.2	5.1	18.4	66	73	288	98
8	21.6	51.8	22.95	30.3	31.66	10.3	35.5	195.9	220.9	219.3	199.0	213.2	25.0	23.4	3.1	17.3	56	60	288	80
9	21.4	51.5	22.39	30.2	31.20	10.1	36.8	192.3	221.3	219.4	199.2	213.4	29.0	27.1	6.9	21.1	88	99	288	114
Average	21.1	51.2	21.92	31.6	32.42	10.7	33.0		221.2	219.5	199.6	212.7								

TABLE VIII

COMPARISON OF AVERAGE COMPOSITION OF VARIOUS VAPOR AND LIQUID STREAMS

	A BENZENE - n-HEPTANE SYSTEM				B n-HEPTANE - TOLUENE SYSTEM				C n-HEXANE - TOLUENE SYSTEM			
	Mole Per Cent Benzene In Primary Condensate				Mole Per Cent n-Heptane in Primary Condensate				Mole Per Cent n-Hexane in Primary Condensate			
	82.5	70.2	54.6	32.6	85.6	63.9	50.1	31.0	80.0	70.2	50.2	21.1
	Mole Per Cent Benzene				Mole Per Cent n-Heptane				Mole Per Cent n-Hexane			
Average y_{A2}	85.10	75.90	63.90	39.70	86.70	67.80	53.00	35.80	85.80	78.70	65.90	31.60
Average y_{A1}	82.57	70.37	54.94	32.88	85.68	64.13	50.32	31.19	80.40	70.93	51.30	21.92
Difference	2.53	5.53	8.96	6.82	1.02	3.67	2.68*	4.61	5.40	7.77	4.60	9.68
Average y_A*	**86.50**	**77.70**	**66.80**	**48.60**	**88.20**	**69.10**	**57.80**	**40.60**	**92.20**	**87.60**	**77.00**	**51.20**
Average y_{A1}	**82.57**	**70.37**	**54.94**	**32.88**	**85.68**	**64.13**	**50.32**	**31.19**	**80.40**	**70.93**	**51.30**	**21.92**
Difference	**3.93**	**7.33**	**11.86**	**15.72**	**2.52**	**4.97**	**7.48**	**9.41**	**11.80**	**16.67**	**25.70**	**29.28**
Average y_A*	**86.50**	**77.70**	**66.80**	**48.60**	**88.20**	**69.10**	**57.80**	**40.60**	**92.20**	**87.60**	**77.00**	**51.20**
Average y_{A2}	**85.10**	**75.90**	**63.90**	**39.70**	**86.70**	**67.80**	**53.00**	**35.80**	**85.80**	**78.70**	**65.90**	**31.60**
Difference	**1.40**	**1.80**	**2.90**	**8.90**	**1.50**	**1.30**	**4.80**	**4.80**	**6.40**	**8.90**	**11.10**	**19.60**
Average y_A*	**86.50**	**77.70**	**66.80**	**48.60**	**88.20**	**69.10**	**57.80**	**40.60**	**92.20**	**87.60**	**77.00**	**51.20**
Average y_{AI}	**85.17**	**76.08**	**64.25**	**40.00**	**86.72**	**67.96**	**53.27**	**36.05**	**86.16**	**79.48**	**67.00**	**32.42**
Difference	**1.33**	**1.62**	**2.55**	**8.60**	**1.48**	**1.14**	**4.53**	**4.55**	**6.04**	**8.12**	**10.00**	**18.78**
Average y_{AI}	**85.17**	**76.08**	**64.25**	**40.00**	**86.72**	**67.96**	**53.27**	**36.05**	**86.16**	**79.48**	**67.00**	**32.42**
Average y_{A1}	**82.57**	**70.37**	**54.94**	**32.88**	**85.68**	**64.13**	**50.32**	**31.19**	**80.40**	**70.93**	**51.30**	**21.92**
Difference	**2.60**	**5.71**	**9.31**	**7.12**	**1.04**	**3.83**	**2.95**	**4.86**	**5.76**	**8.55**	**15.70**	**10.50**
Average y_{AI}	**85.17**	**76.08**	**64.25**	**40.00**	**86.72**	**67.96**	**53.27**	**36.05**	**86.16**	**79.48**	**67.00**	**32.42**
Average y_{A2}	**85.10**	**75.90**	**63.90**	**39.70**	**86.70**	**67.80**	**53.00**	**35.80**	**85.80**	**78.70**	**65.90**	**31.60**
Difference	**0.07**	**0.18**	**0.35**	**0.30**	**0.02**	**0.16**	**0.27**	**0.25**	**0.36**	**0.78**	**1.10**	**0.82**
Average x_A	82.50	70.20	54.60	32.60	85.60	63.90	50.10	31.00	80.00	70.20	50.20	21.10
Average x^*_{AI}	81.20	67.70	51.20	24.90	83.80	62.50	44.70	26.30	67.50	54.50	35.80	10.70
Difference	1.30	2.50	3.40	7.70	1.80	1.40	5.40	4.70	12.50	15.70	14.40	10.40

interfacial temperature and bubble point of condensate is smaller than difference between the dew point and boiling point of the condensate. At concentrations greater than 50 mole per cent of the more volatile component, for both benzene-n-heptane and n-heptane-toluene systems, the difference between the interface and the bubble point temperature is 1°F. or less, whereas for n-hexane-toluene system the minimum difference in the two temperatures is 6°F at 80 mole per cent n-hexane in the condensate. Referring to Table VIII, one notices that at concentrations greater than 50 mole per cent of the more volatile components, for both benzene-n-heptane and n-heptane-toluene systems the difference in composition of the condensate from vapor-liquid interface to the condenser surface is less than 5 mole per cent, compared to 15 mole per cent difference for n-hexane-toluene system. This large change in composition across the liquid-film for n-hexane-toluene system corresponds to the wide difference which exists between the interfacial temperature and the boiling point of the condensate for this system. Therefore, where there is negligible or small composition change across the liquid-film, the boiling point of the condensate may be taken as the interfacial temperature. However, where there is significant change in composition across the liquid-film, the boiling point at the vapor-liquid interface could be significantly different from that based on the average condensate composition; that is, the boiling point of condensate will not be the same as the interfacial temperature.

The calculated film coefficient of heat transfer are tabulated in Table VII-A, B, and C and are presented graphically as a function of

TABLE IX

COMPARISON OF INTERFACIAL TEMPERATURE WITH VAPOR TEMPERATURE, DEW POINT, AND BUBBLE POINT OF PRIMARY CONDENSATE

	A BENZENE - n-HEPTANE SYSTEM				B n-HEPTANE - TOLUENE SYSTEM				C n-HEXANE - TOLUENE SYSTEM			
	Mole Per Cent Benzene in Primary Condensate				Mole Per Cent n-Heptane in Primary Condensate				Mole Per Cent n-Hexane in Primary Condensate			
	82.5	70.2	54.6	32.6	85.6	63.9	50.1	31.0	80.0	70.2	50.2	21.1
	TEMPERATURE, °F.				TEMPERATURE, °F.				TEMPERATURE, °F.			
Average T_V	179.8	184.4	190.8	198.7	210.9	213.8	216.9	222.5	175.9	185.1	201.2	221.2
Average T_I	178.4	180.9	185.0	194.8	210.3	212.8	215.9	220.3	168.6	175.1	186.5	212.7
Difference	1.4	3.5	5.8	3.9	0.6	1.0	1.0	2.2	7.3	10.0	14.7	8.5
Average T_{Dp}	179.0	182.9	188.8	198.0	210.4	213.6	216.6	221.6	174.8	183.5	200.0	219.5
Average T_I	178.4	180.9	185.0	194.8	210.3	212.8	215.9	220.3	168.6	175.1	186.5	212.7
Difference	0.6	2.0	3.8	3.2	0.1	0.8	0.7	1.3	6.2	8.4	13.5	6.8
Average T_I	178.4	180.9	185.0	194.8	210.3	212.8	215.9	220.3	168.6	175.1	186.5	212.7
Average T_{Bp}	178.2	180.3	184.1	191.2	210.3	212.5	214.9	218.9	162.9	167.4	177.3	199.6
Difference	0.2	0.6	0.9	3.6	0.0	0.3	1.0	1.4	5.7	7.7	9.2	13.1

temperature drop across the film in Figure 18 through 29.

At high concentration of benzene in n-heptane, where vapor temperature, dew point and bubble point of condensate, and the interfacial temperature are very close together, the heat transfer coefficients based on these temperatures are almost identical and all fall between the coefficients for pure benzene and pure n-heptane. However, as the concentration of the more volatile components, benzene, is reduced the heat transfer data based on vapor temperature, dew point of the condensate and the interfacial temperature become smaller than those based on bubble point temperature and gradually fall below those of n-heptane. It is also noticeable from Figures 18 through 21 that as the concentration of benzene in mixtures is reduced the curves move away from the curve for pure benzene and gradually get closer to that of pure n-heptane. This is consistent with the assumption that heat transfer coefficient, like any other property of ideal mixtures, is a weighted average of those for the pure materials.

At high concentration of benzene the interfacial heat transfer coefficient is very close to that based on the bubble point of condensate but, as the benzene concentration is reduced the interfacial heat transfer coefficient moves toward the heat transfer data based on the dew point of the condensate. Again this is consistent with the observation that as concentration of n-heptane, the less volatile component, in mixture is increased, the interfacial temperature gradually becomes closer to the dew point of the condensate than to its boiling point, (See Table IX). Furthermore, mass transfer data show that the gas-film resistance constitutes a much smaller percentage of the total resistance at high

concentration of n-heptane than at low concentration. This seems to indicate that at low concentration of the more volatile component, the dew point could determine the temperature driving force across the condensate film.

At high concentration of n-heptane in toluene, the curves of film coefficients calculated with the four temperature differences fall below those of pure components. As the concentration of n-heptane in the mixture is reduced, the curves move upward and finally at 31 mole per cent n-heptane the curves fall above that of pure toluene. Referring to Table IX, it is noticed that for each mixture of n-heptane-toluene system, the interfacial temperature, vapor temperature, dew point and the boiling temperatues are very close to each other, which explains why the film coefficient curves are so close together. Figures 22 through 25 show that for all concentrations of n-heptane, the coefficient decrease with increasing temperature drop across the film up to a certain degree, just as did the pure components comprising the mixture. Then, with further increase in temperature drop, the coefficients again rise. This effect is already explained, for pure components, as being the transition point where the flow of the condensate changes from streamline flow to turbulent.

In all concentrations of n-hexane in toluene the heat transfer data based on vapor temperature and the dew point of condensate are very close together and all fall below those for both of the pure components. This is expected, because for all systems the vapor and the dew point temperatures are very close together. The heat transfer data based on

interfacial temperature fall between those based on the dew point and the boiling point of condensate. Only at high concentration of one of the components in the mixture, the heat transfer data based on boiling point of the condensate fall between those of the pure components. In intermediate concentrations the data fell above those of the pure components.

One may expect that the heat transfer data for n-hexane-toluene system would follow the same pattern as the data for n-heptane-toluene system. However, the plotted data in Figures 26 through 29 show otherwise. A possible reason for this deviation may be offered here.

At high concentration of n-hexane in toluene difficulty was encountered in maintaining a uniform smooth film of condensate. At concentration of 80.0 mole per cent n-hexane it was observed when the vapor first entered the jacket, and for about 10 minutes thereafter, a smooth film similar to that of the other systems formed. Shortly thereafter the flow of condensate appeared to be disturbed and a ring formation of the condensate appeared on the condensing surface, Figure 30. These rings of condensate, which were moving along the tube in the direction of vapor flow, were uniformly spaced and were about 1/8 to 3/16 inches in width and extended completely around the tube. There were four rings per inch of tube length. The boundaries of the rings were very well defined. However, as the temperature difference across the film was reduced the shape of the rings became irregular and they were no longer uniformly spaced. The flow of the condensate became more irregular as the temperature drop was lowered further. Now ridges and valleys with definite boundaries were taking the place of uniform rings, and very often they would run

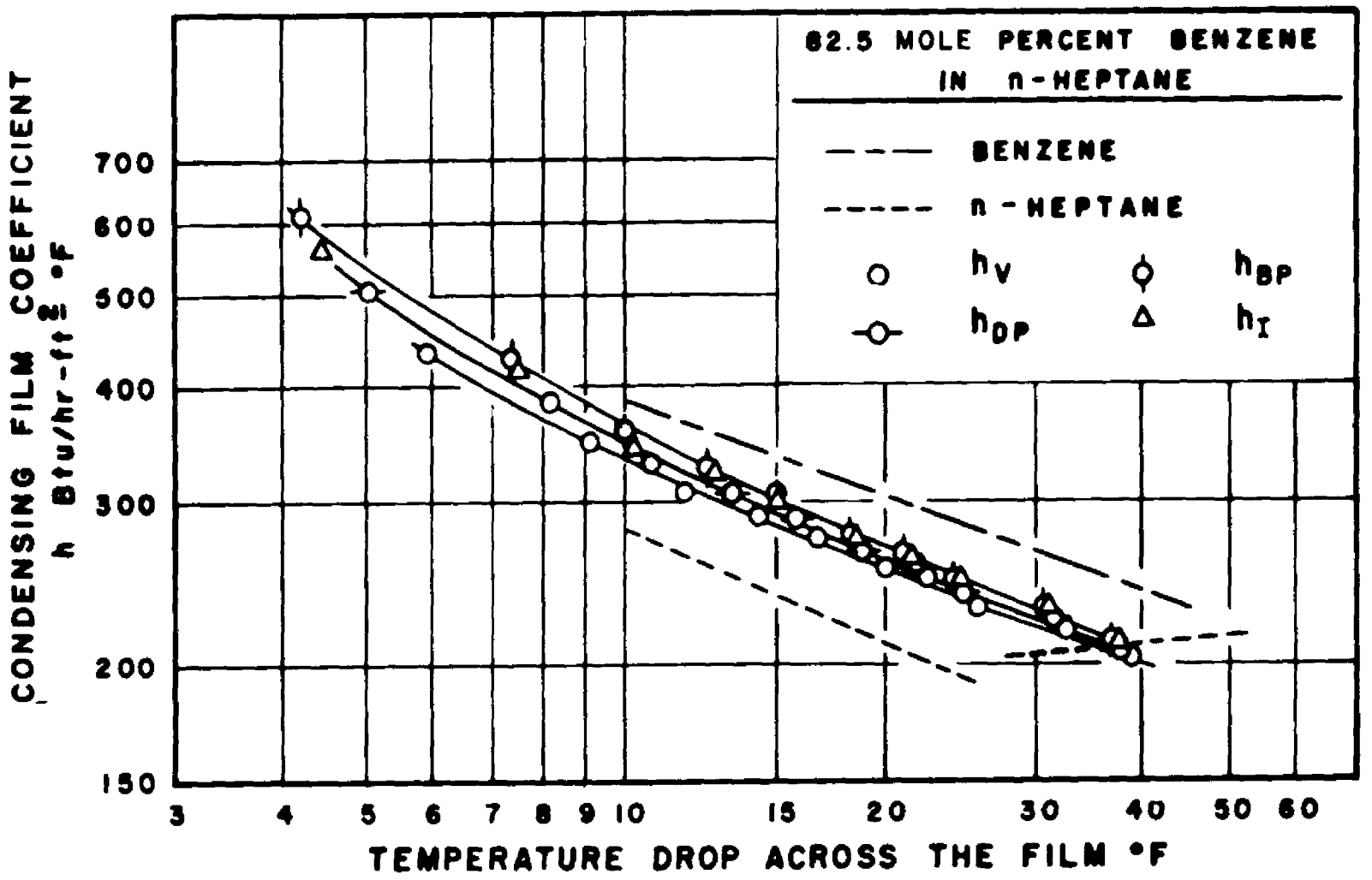


FIGURE 18 CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR BENZENE — n-HEPTANE SYSTEM

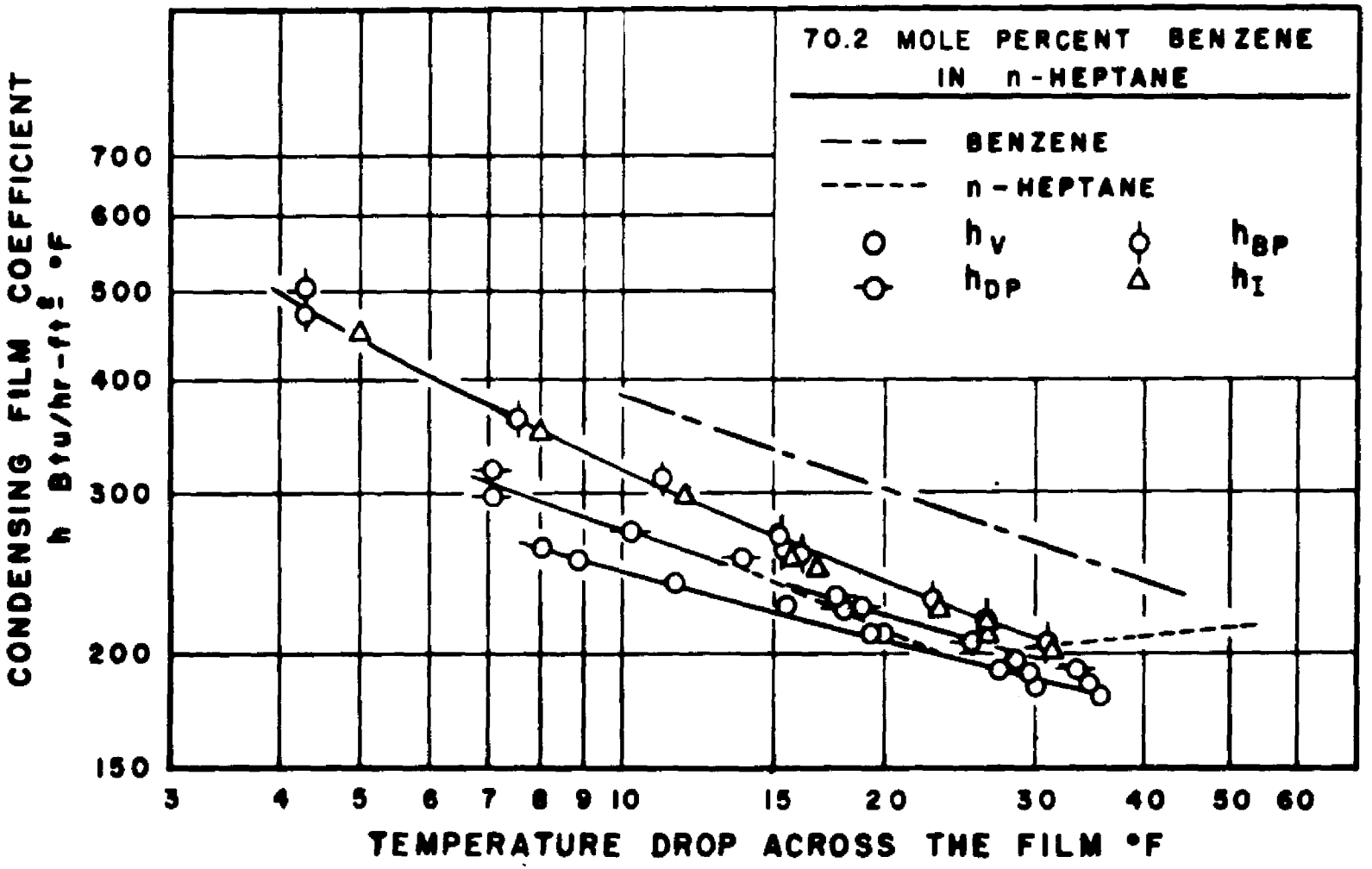


FIGURE 19. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR BENZENE–n-HEPTANE SYSTEM

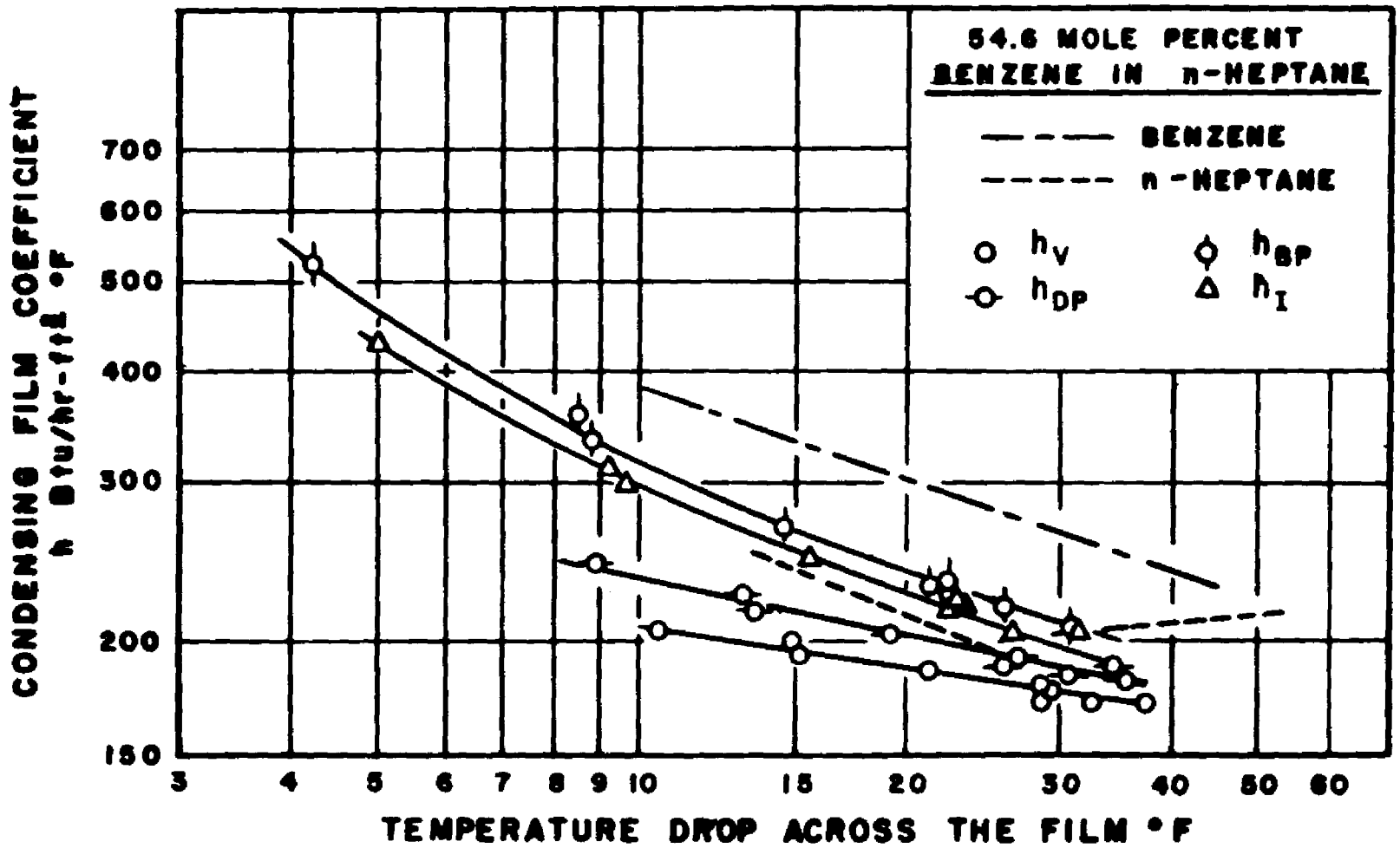


FIGURE 20. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR BENZENE–n-HEPTANE SYSTEM

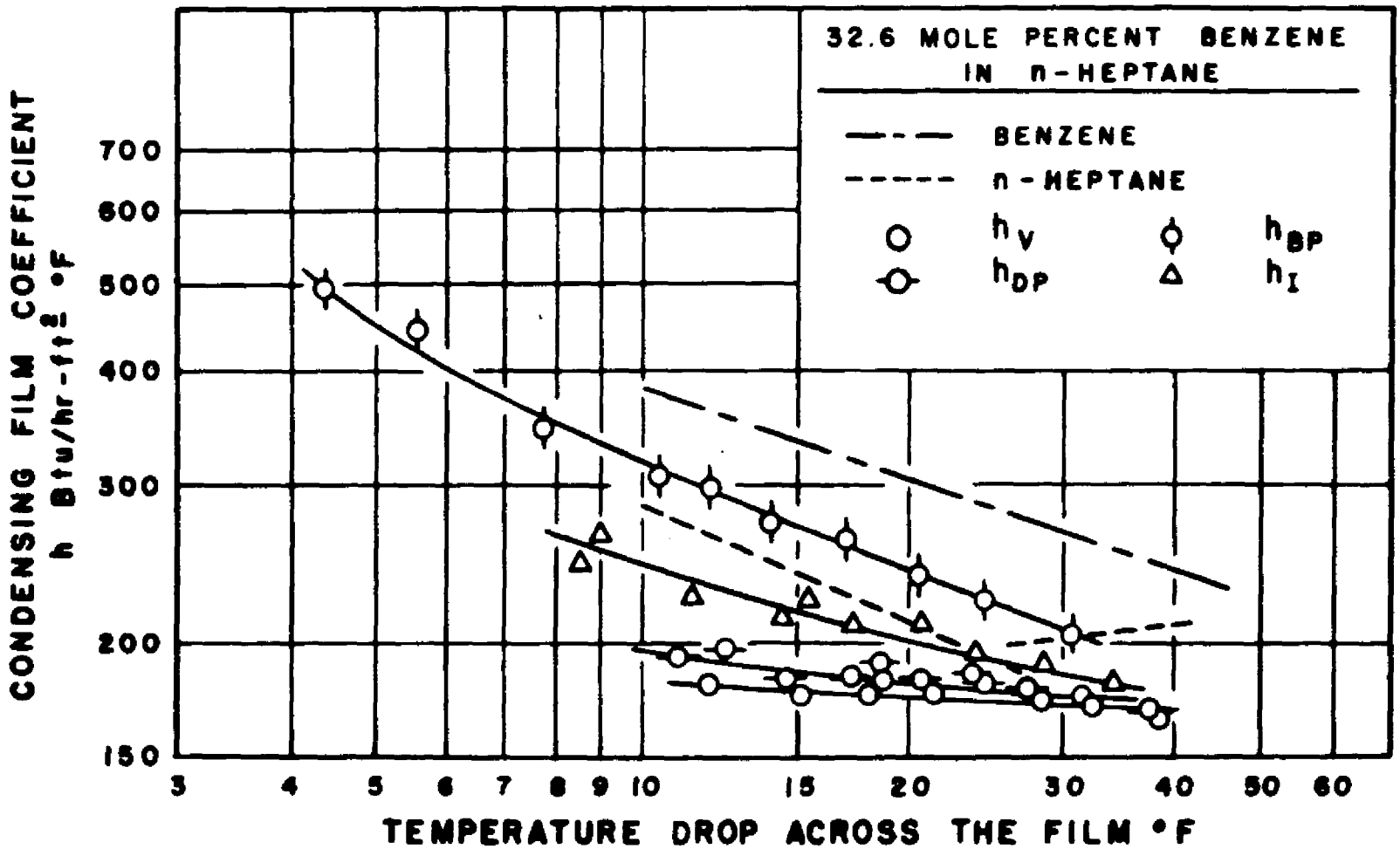


FIGURE 21. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR BENZENE – n-HEPTANE SYSTEM

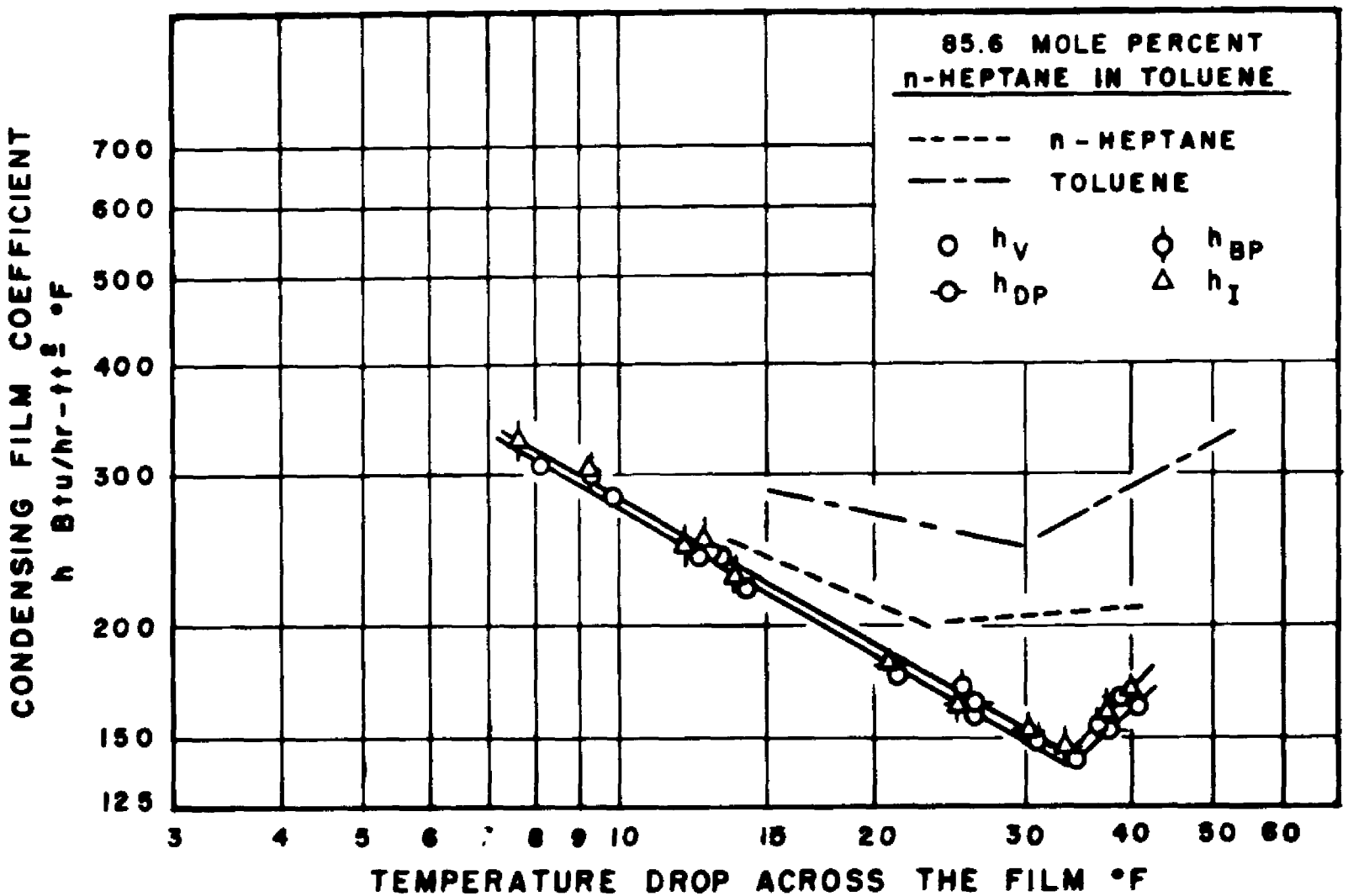


FIGURE 22. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEPTANE – TOLUENE SYSTEM

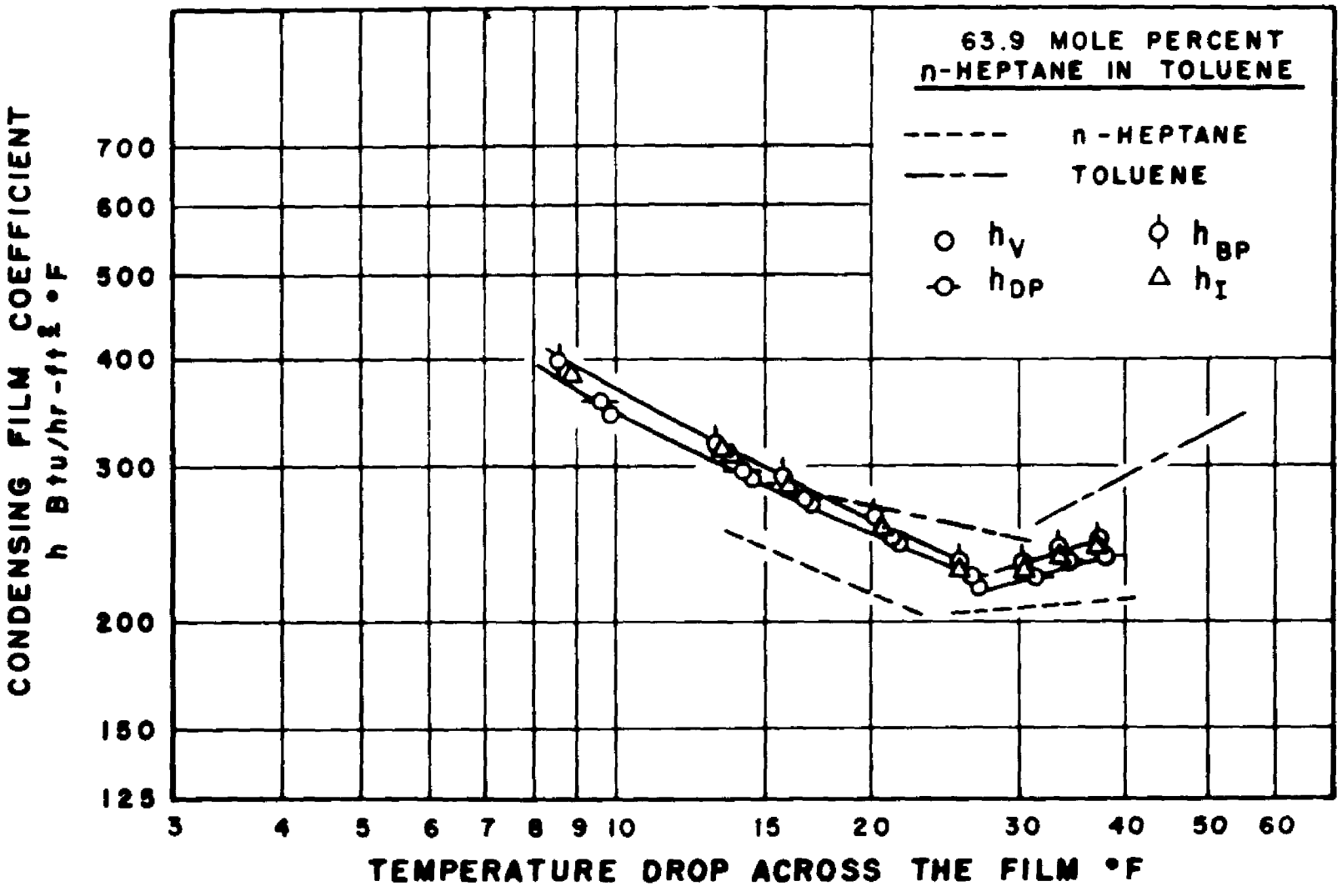


FIGURE 23. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEPTANE - TOLUENE SYSTEM

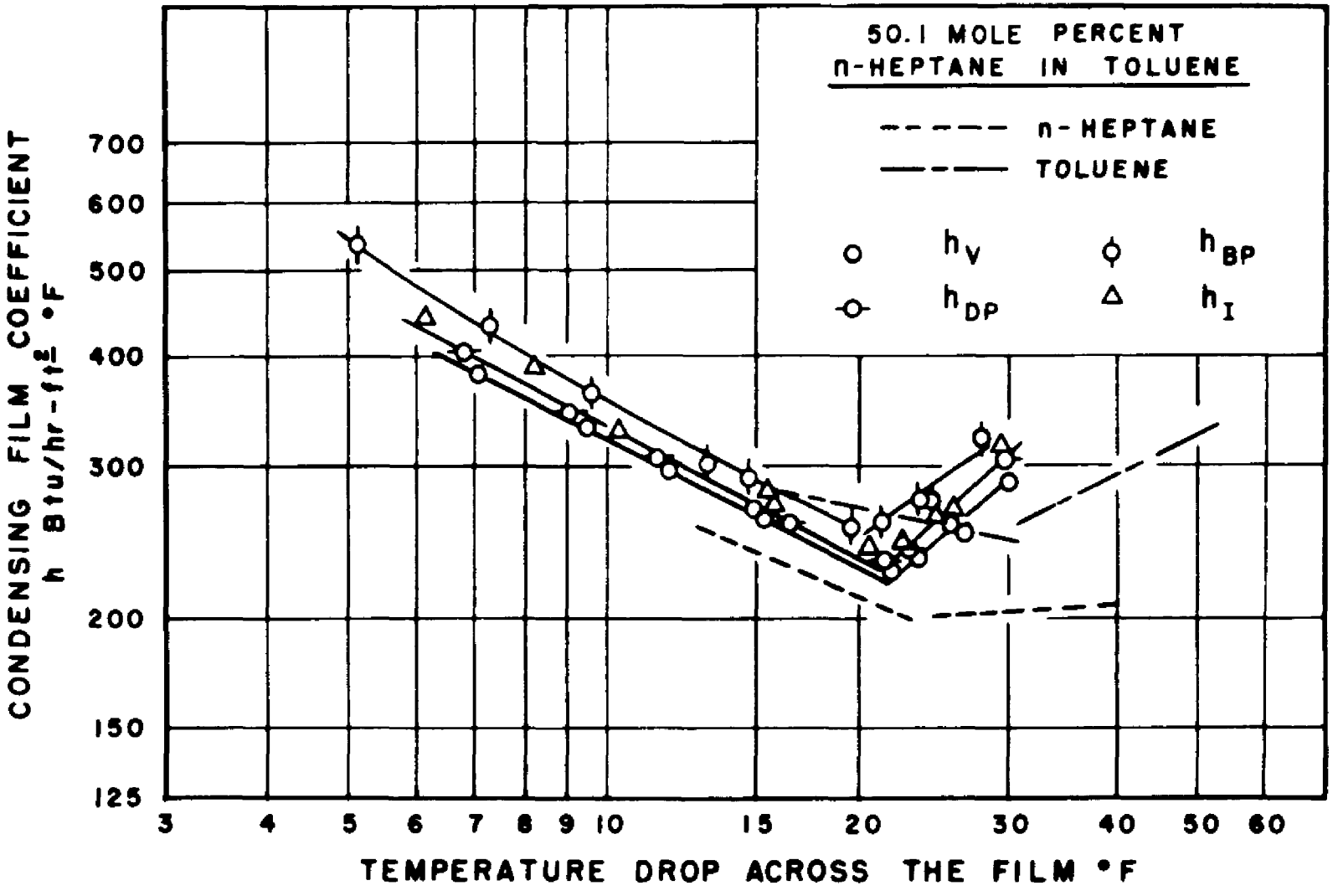


FIGURE 24. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEPTANE – TOLUENE SYSTEM

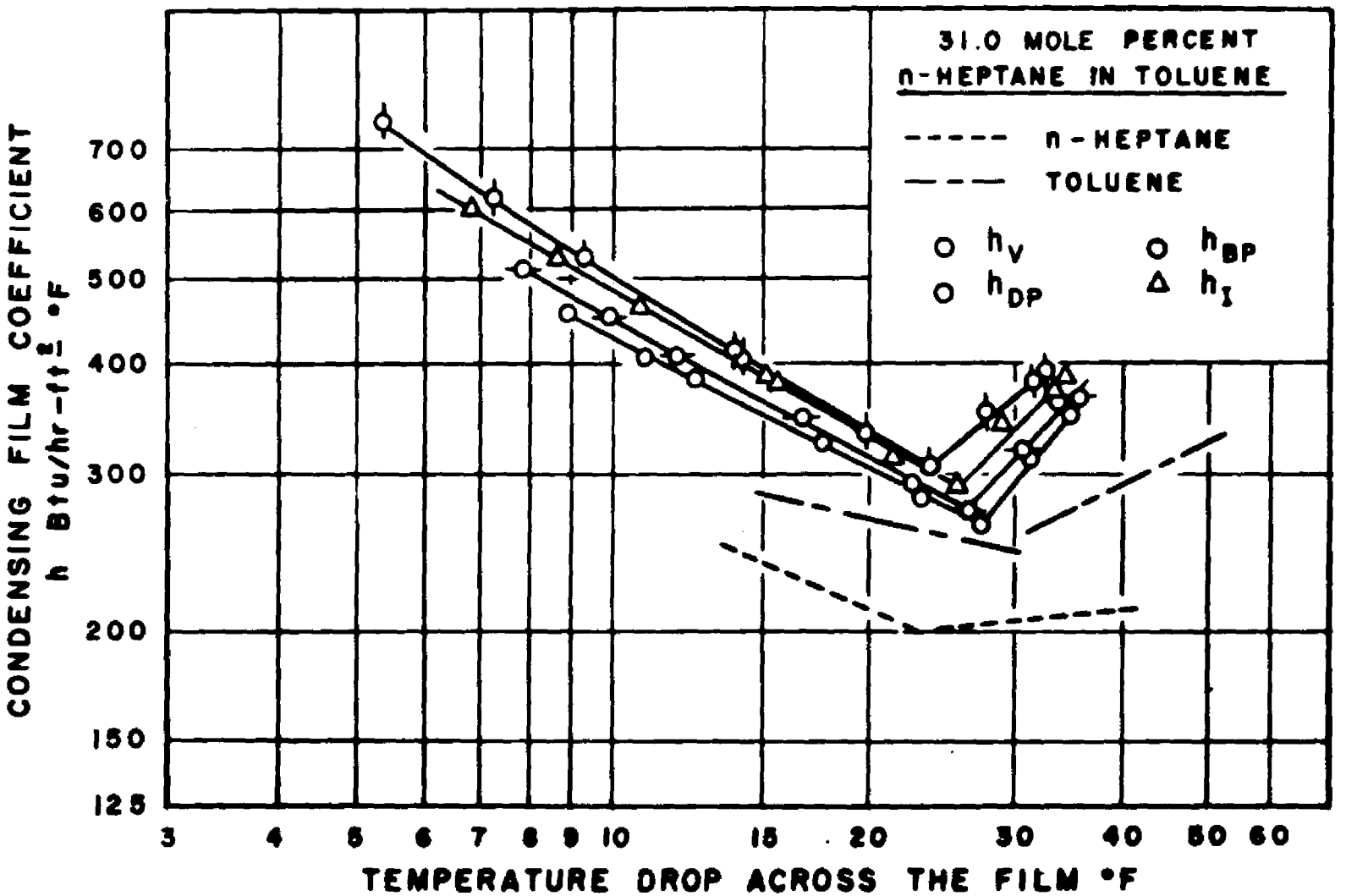


FIGURE 25. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEPTANE - TOLUENE SYSTEM

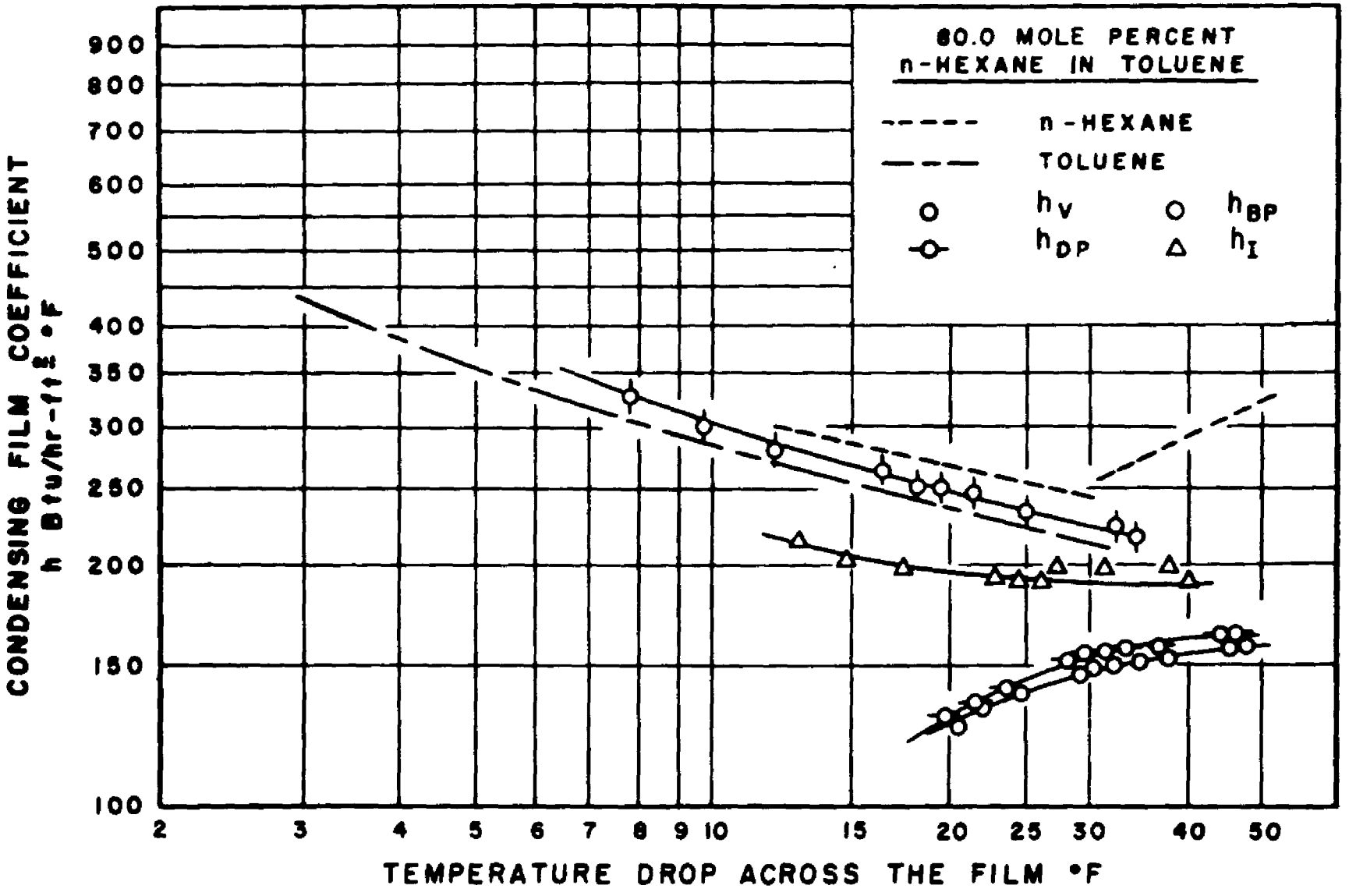


FIGURE 26. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEXANE – TOLUENE SYSTEM

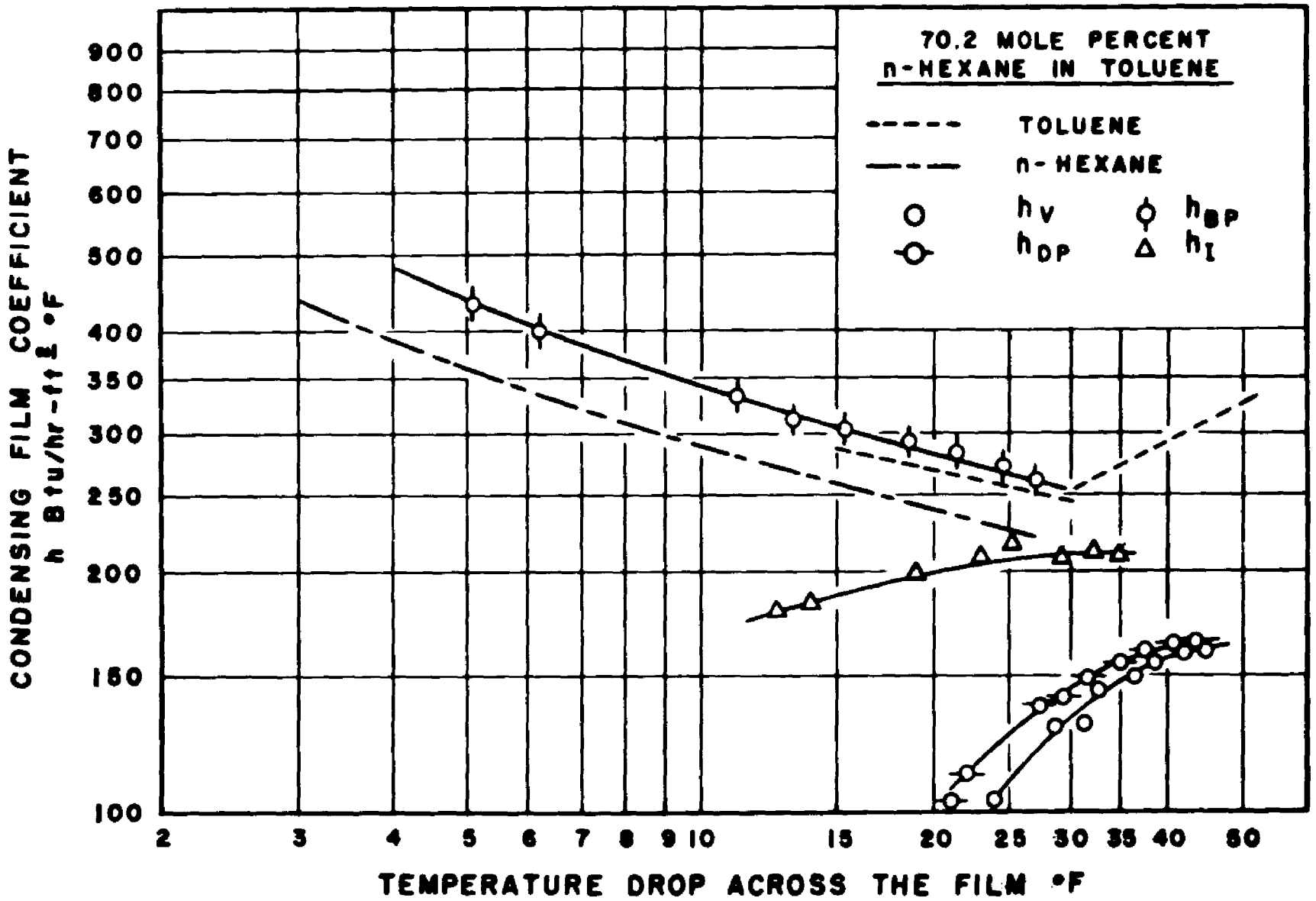


FIGURE 27. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEXANE – TOLUENE SYSTEM

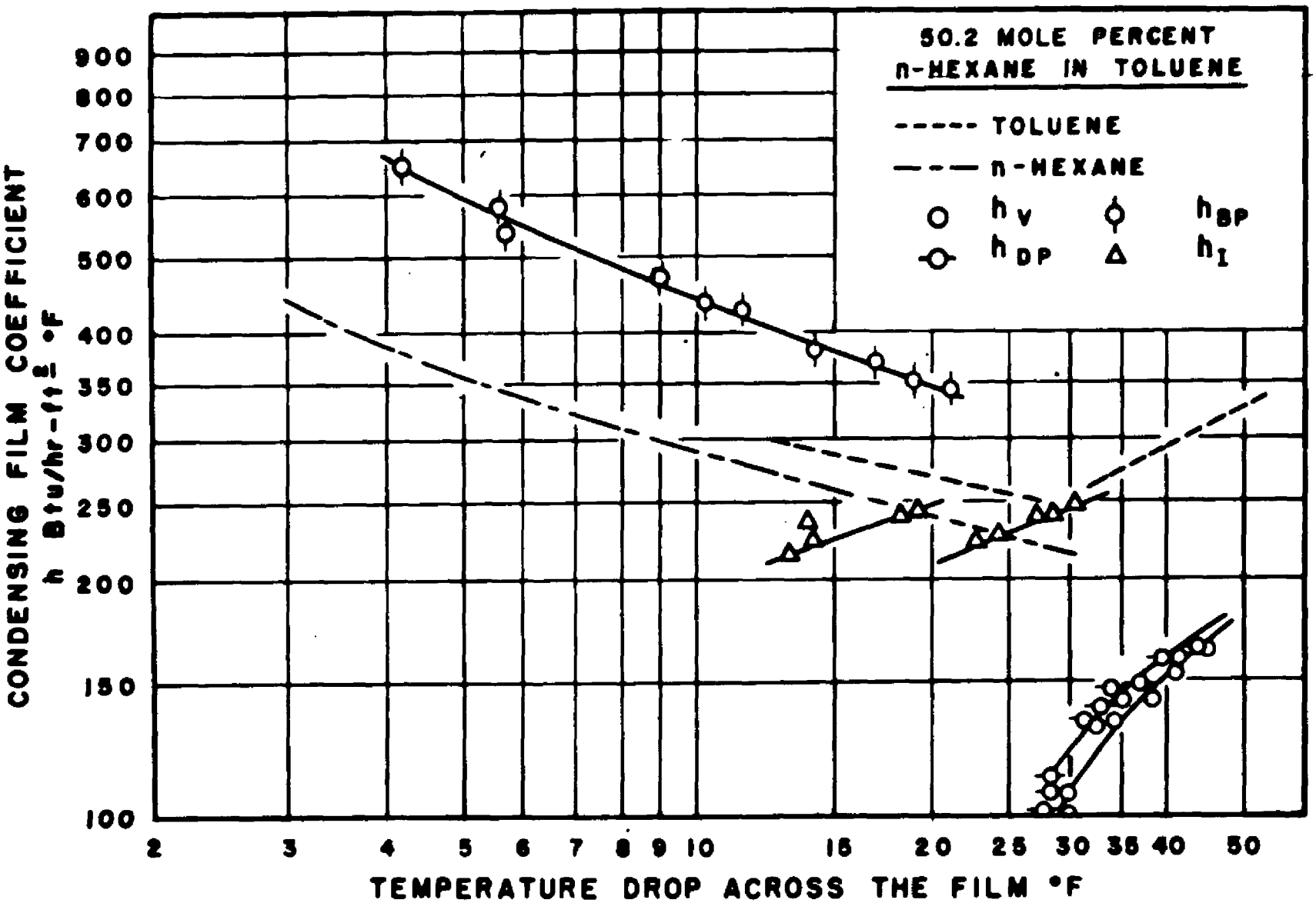


FIGURE 28. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEXANE – TOLUENE SYSTEM

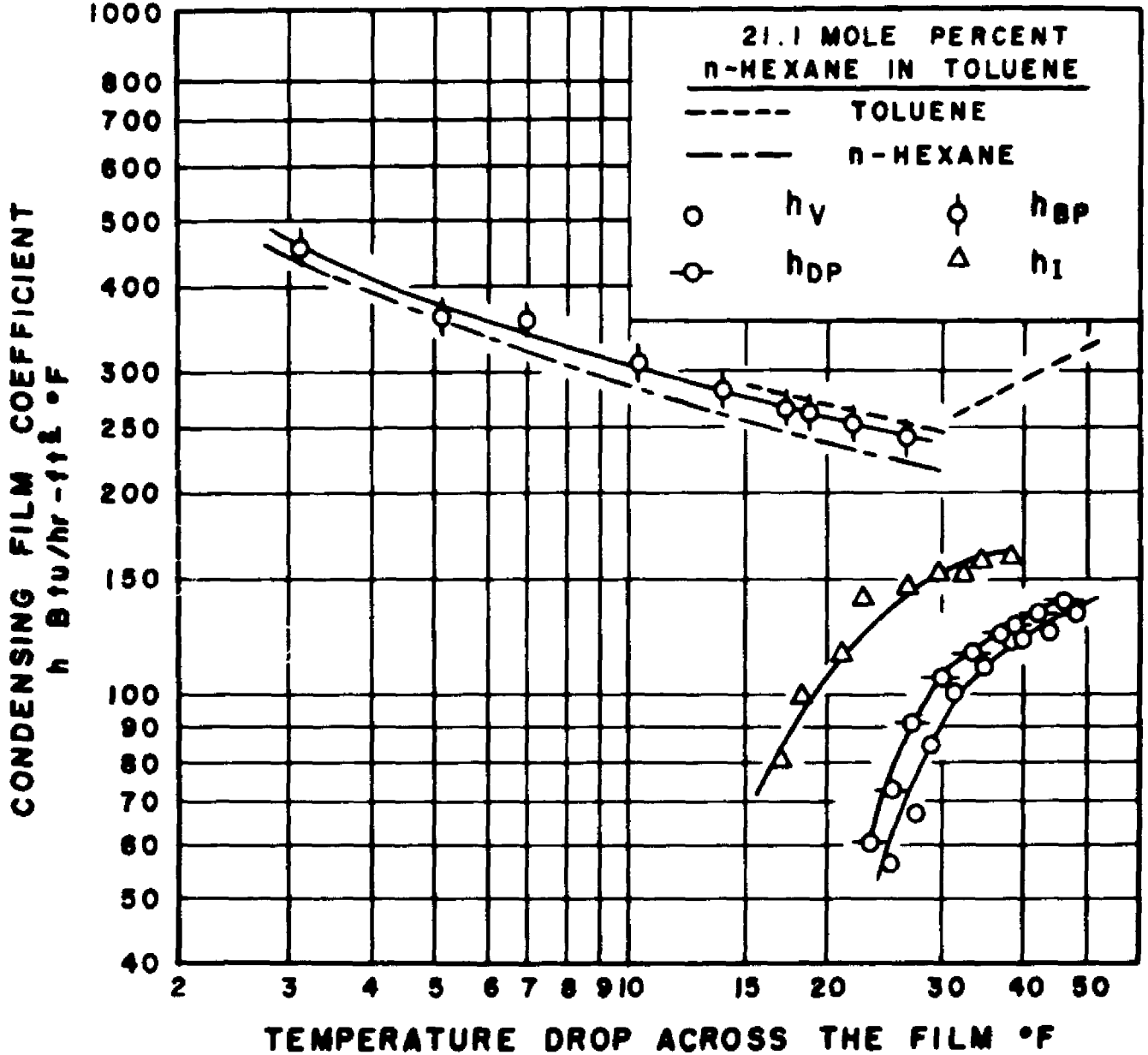


FIGURE 29. CORRELATION OF CONDENSING FILM COEFFICIENT WITH TEMPERATURE DIFFERENCE FOR n-HEXANE - TOLUENE SYSTEM

into each other. The valleys appeared to be very thin film of the condensate, which in places showed disturbances because of the flow of large ridges. These irregular flow patterns indicate turbulence in the condensate film and this could be true even though the average condensate rate per unit length of the tube indicates a Reynolds number well below the accepted critical value of 4200, if this is the true criterion.

The irregular pattern of flow, Figure 31, was more pronounced when the concnetration of n-hexane in toluene was lowered to 50 per cent and the temperature drop and rate of condensation were maintained in the intermediate range. As the rate of condensation was further reduced, the flow pattern was extremely non-uniform and the valleys were becoming wider and wider. At times it seemed as though there were no flow of the condensate on the condensing bar except for a few ridges at a distance of about 1/2 to 3/4 inches apart. It is believed that this flow pattern accounts for the high coefficients determined for 50.2 mole per cent of n-hexane in toluene. A further reduction in concentration of n-hexane in toluene, to 21.1 mole per cent, resulted in uniform ring formation only at a high rate of condensation. For both 50.2 mole per cent and 21.1 mole per cent n-hexane, the curves based on temperature difference between the entering vapor and average wall temperature are not straight line and turn upward at low temperature differences. This is explained by the formation of ridges and the large gap between them.

The ring and irregular ridge formation for the system of n-hexane and toluene can be summarized as follows: At either high, (β0 mole per

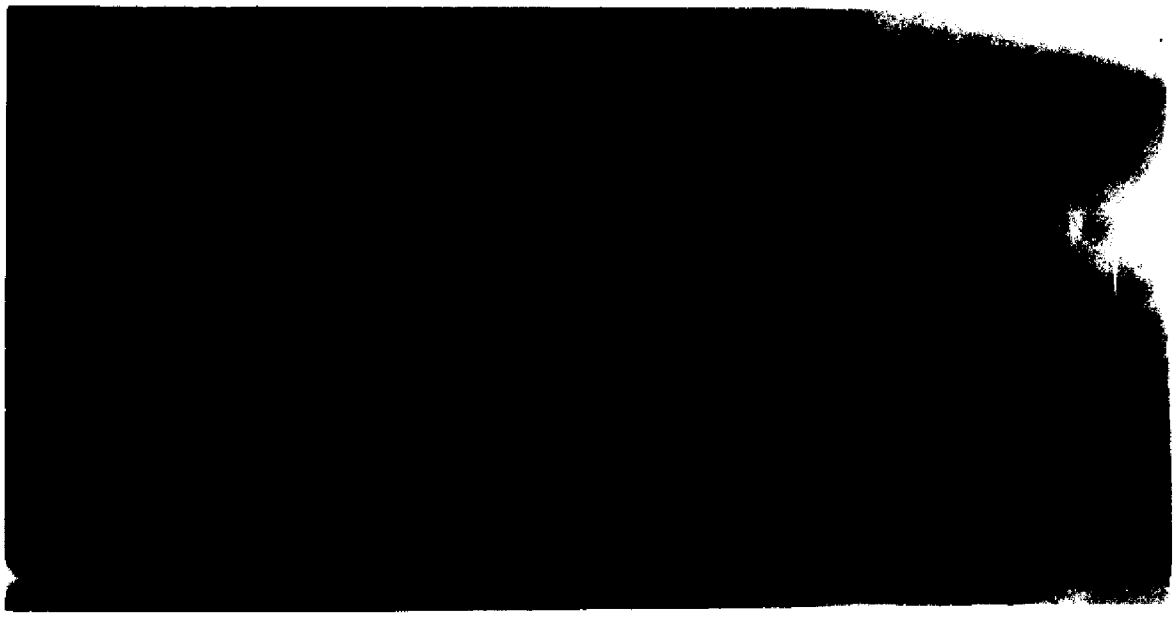


FIGURE 30. RING FORMATION OF THE CONDENSATE AT HIGH CONCENTRATION OF n-HEXANE IN TOLUENE, INTERMEDIATE CONDENSATE RATE, AND INTERMEDIATE TEMPERATURE DIFFERENCE.

FIGURE 31. IRREGULAR PATTERN OF CONDENSATE AT LOW CONCENTRATION OF n-HEXANE IN TOLUENE, LOW CONDENSATE RATE, AND INTERMEDIATE TEMPERATURE DIFFERENCE.

cent) or low, (21.1 mole per cent) concentration of n-hexane in toluene regular ring formations were produced at high condensation rates. Such rings remained as the condensation rate was lowered to a medium range. However, as the condensation rate was reduced further, irregular ridges took the place of uniform rings. At a lower concentration, a 50 per cent mixture, formation of the irregular ridges took place much faster and sooner and as the condensation rate was lowered the flow patterns became worse.

This same sort of film appearance has been reported by Bromley, et al. (9) for condensing pure water and pure acetone on the outside of a 3/4 inch I.D. horizontal stainless steel tube. Todd (104) observed the same flow pattern of the condensate for a mixture of methanol in benzene, but the flow of irregular ridges were not as severe as in the present work. The ring formation and rippling of condensate would increase the rate of condensation, as compared to a uniformly distributed film. The increase in rate of condensation with the consequent turbulence in the liquid film would result in high heat transfer coefficient.

From plots of heat transfer coefficients and temperature drop across the film for each system, Figures 18 through 29, the heat transfer coefficients at 5°F. intervals of temperature drop were recorded and tabulated in Table X -A, B, and C. The heat transfer data based on vapor temperature were eliminated. It has been pointed out previously, that in all cases the vapor temperature and the dew point of the condensate were close to each other. Consequently heat transfer

data were close together. Therefore, it is safe to assume that, for purposes of cross-correlation the behavior of heat transfer data based on dew point temperature would also represent the behavior of data based on vapor temperature. The data in Table X were plotted on Figures 32, 33, and 34. These figures show, for each of the three methods of measuring temperature drop across the film, the relationship of heat transfer coefficient as a function of condensate composition with tempearture drop across the film as a parameter.

Comparison of the three sets of curves for each of the three binary systems shows that only for the benzene-n-heptane system were the curves obtained based on boiling point and the interfacial temperatures most linear; even they were not completely so. These curves are more nearly linear at the higher temperature differences. The maximum deviation from linearity for curves based on boiling point temperature was -8 per cent against the maximum deviation of -12 per cent obtained for the curves based on interfacial temperature. Both of these deviations were obtained for a 10°F. temperature drop across the film. The deviations become much smaller at higher temperature differences.

The curves based on dew point of condensate show a large dip at approximately 33.0 mole per cent benzene in n-heptane. Reviewing the temperature-composition diagram for this system, it is noticed that the temperature difference between the dew point and the bubble point of condensate is largest at a composition of 30 to 35 mole per cent than at any other composition. Moreover, a study of Figures 20 and 21 shows that at compositions of 54.6 and 32.6 mole per cents the heat transfer

TABLE X

FILM COEFFICIENT OF CONDENSING BINARY VAPORS AT

CONSTANT TEMPERATURE DROP

A BENZENE-n-HEPTANE SYSTEM

Temperature Difference °F.	Mole Per Cent Benzene in the Primary Condensate					
	0	32.6	54.6	70.2	82.5	100
$\Delta T_I = T_I - T_W$	h_I, Btu/Hr. Ft.^{2} °F.					
10	280	258	300	310	350	380
15	237	227	258	265	298	333
20	210	210	230	238	263	300
25	192	196	212	220	240	280
30	205	187	198	204	223	264
$\Delta T_{DP}=T_{DP} - T_W$	h_{DP}, Btu/Hr. Ft.^{2} °F.					
10	280	197	235	271	347	380
15	237	186	215	240	290	333
20	210	181	203	222	260	300
25	192	177	195	208	238	280
30	205	175	187	197	222	264
35	209	173	181	189	210	250
$\Delta T_{BP}=T_{BP} - T_W$	h_{BP}, Btu/Hr. Ft.^{2} °F.					
10	280	316	315	320	360	380
15	237	268	268	272	305	333
20	210	240	240	243	268	300
25	192	222	222	223	245	280
30	205	206	208	207	237	264
35	209	195	200	195	215	250

TABLE X

FILM COEFFICIENT OF CONDENSING BINARY VAPORS AT

CONSTANT TEMPERATURE DROP

B n-HEPTANE-TOLUENE SYSTEM

Temperature Difference °F.	Mole Per Cent n-Heptane in the Primary Condensate					
	0	31.0	50.1	63.9	85.6	100.0
$\Delta T_I = T_I - T_W$	h_I, Btu/Hr. Ft.^{2} °F.					
10	312	485	345	360	280	285
15	285	390	281	295	225	240
20	268	330	245	255	192	212
25	255	293	272	228	172	202
30	245	341	320	228	155	205
$\Delta T_{DP} = T_{DP} - T_W$	h_{DP}, Btu/Hr. Ft.^{2} °F.					
10	312	450	333	348	280	285
15	285	365	270	285	223	240
20	268	315	233	252	190	212
25	255	280	262	228	177	202
30	245	310	305	220	152	205
35	275	365	335	230	147	208
$\Delta T_{BP} = T_{BP} - T_W$	h_{BP}, Btu/Hr. Ft.^{2} °F.					
10	312	505	357	368	280	285
15	285	398	290	300	223	240
20	268	338	250	260	190	212
25	255	320	287	232	177	202
30	245	368	320	230	152	205
35	275	410	355	240	147	208

TABLE X

FILM COEFFICIENT OF CONDENSING BINARY VAPORS AT

CONSTANT TEMPERATURE DROP

C n-HEXANE-TOLUENE SYSTEM

Temperature Difference °F.	Mole Per Cent n-Hexane in the Primary Condensate					
	0	21.1	50.2	70.2	80.0	100.0
$\Delta T_I = T_I - T_W$	h_I, Btu/Hr. Ft.^{2} °F.					
15	285	-	220	182	205	257
20	268	104	228	198	197	238
25	255	132	232	207	193	225
30	245	150	240	210	190	214
35	270	158	247	210	190	-
$\Delta T_{DP} = T_{DP} - T_W$	h_{DP}, Btu/Hr. Ft.^{2} °F.					
25	255	74	92	125	144	225
30	245	104	120	142	155	214
35	270	120	142	154	160	-
40	290	130	158	162	163	-
45	305	137	173	165	164	-
$\Delta T_{BP} = T_{BP} - T_W$	h_{BP}, Btu/Hr. Ft.^{2} °F.					
10	312	300	445	342	300	288
15	286	272	382	303	265	257
20	268	255	345	281	247	238
25	255	243	320	265	233	225
30	245	235	300	252	223	214

coefficient curves based on dew point temperature fall below those of the pure components. These can account for the large dip for curves based on dew point tempearture.

For the n-heptane-toluene system the curves based on the three different temperature differences are quite similar in appearance as would be expected since there is only a small difference between the interfacial temperature, dew point and boiling point temperatures. It is also noticed that for this binary system a sharp peak at 31.0 mole per cent and a dip at 85.6 mole per cent occur in values of heat transfer coefficients. Examination of the temperature-composition diagram for the system shows that the composition at which the peak occurs is in the region that the difference between the boiling and dew point is the greatest. Furthermore, reviewing Figures 22 through 25, one would notice that at 31.0 mole per cent n-heptane, Figure 25, the heat transfer coefficient curves all fall above those of the pure components. However, the curves for 50.1 and 63.9 mole per cents n-heptane fall between those of the pure hydrocarbons. Finally, those of 85.6 mole per cent all fall below the curves for the pure components. This explains the dip which is present at 85.6 mole per cent n-heptane in Figure 33. Also, Figure 33 shows that the curves at temperature differences of 30°F. do not fall below those of 25°F lines up to about 70 mole per cent n-heptane. A quick check of Figures 22 to 25 would reveal the reason is that at 31.0, 50.1, and 63.9 mole per cents n-heptane film coefficients of heat transfer at 30°F. temperature difference are higher than those at 25°F. temperature drop. Only at 85.6

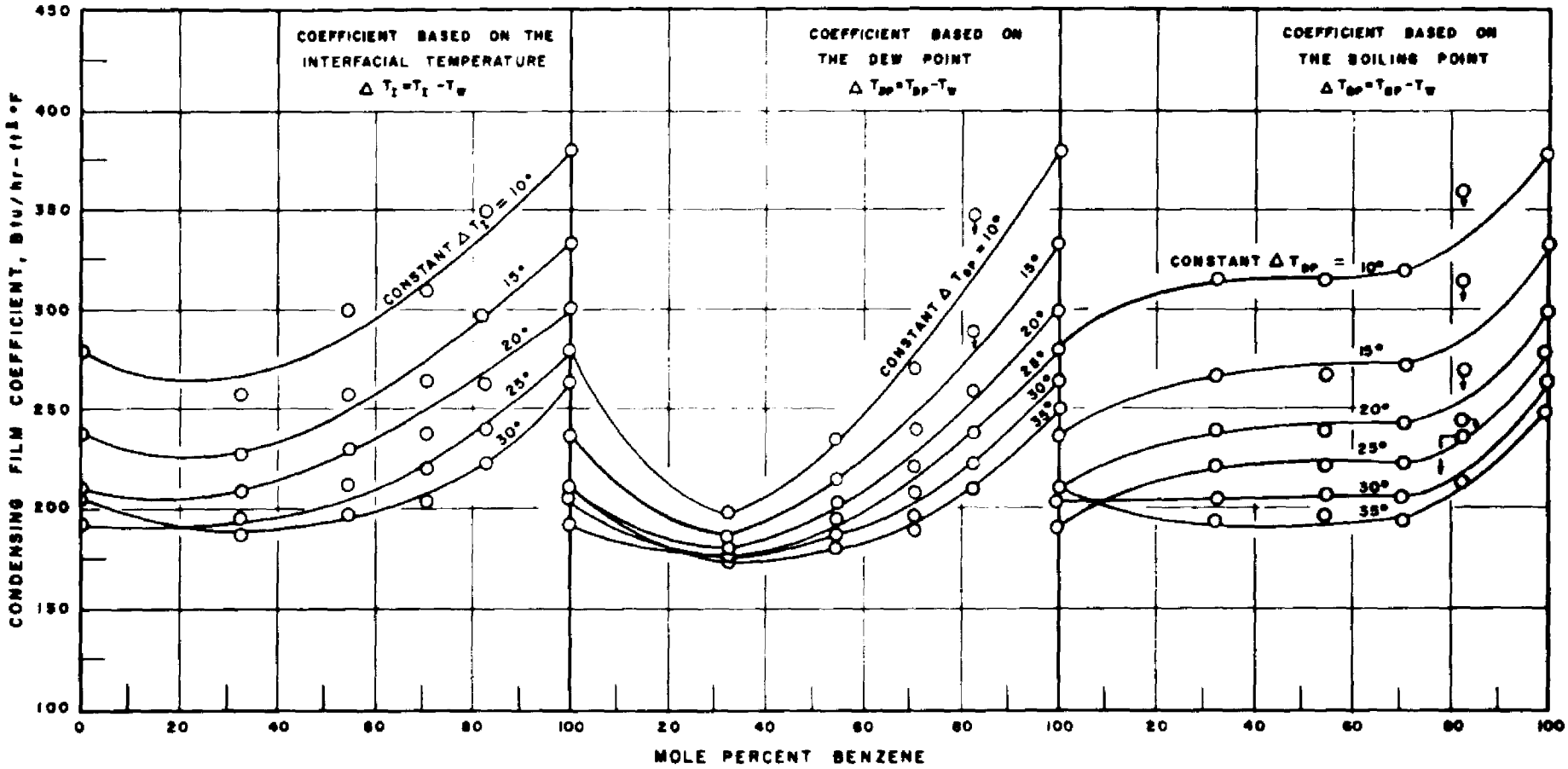


FIGURE 32. VARIATION OF CONDENSING FILM COEFFICIENT WITH COMPOSITION OF BINARY MIXTURES AT CONSTANT TEMPERATURE DIFFERENCE
(BENZENE – n-HEPTANE SYSTEM)

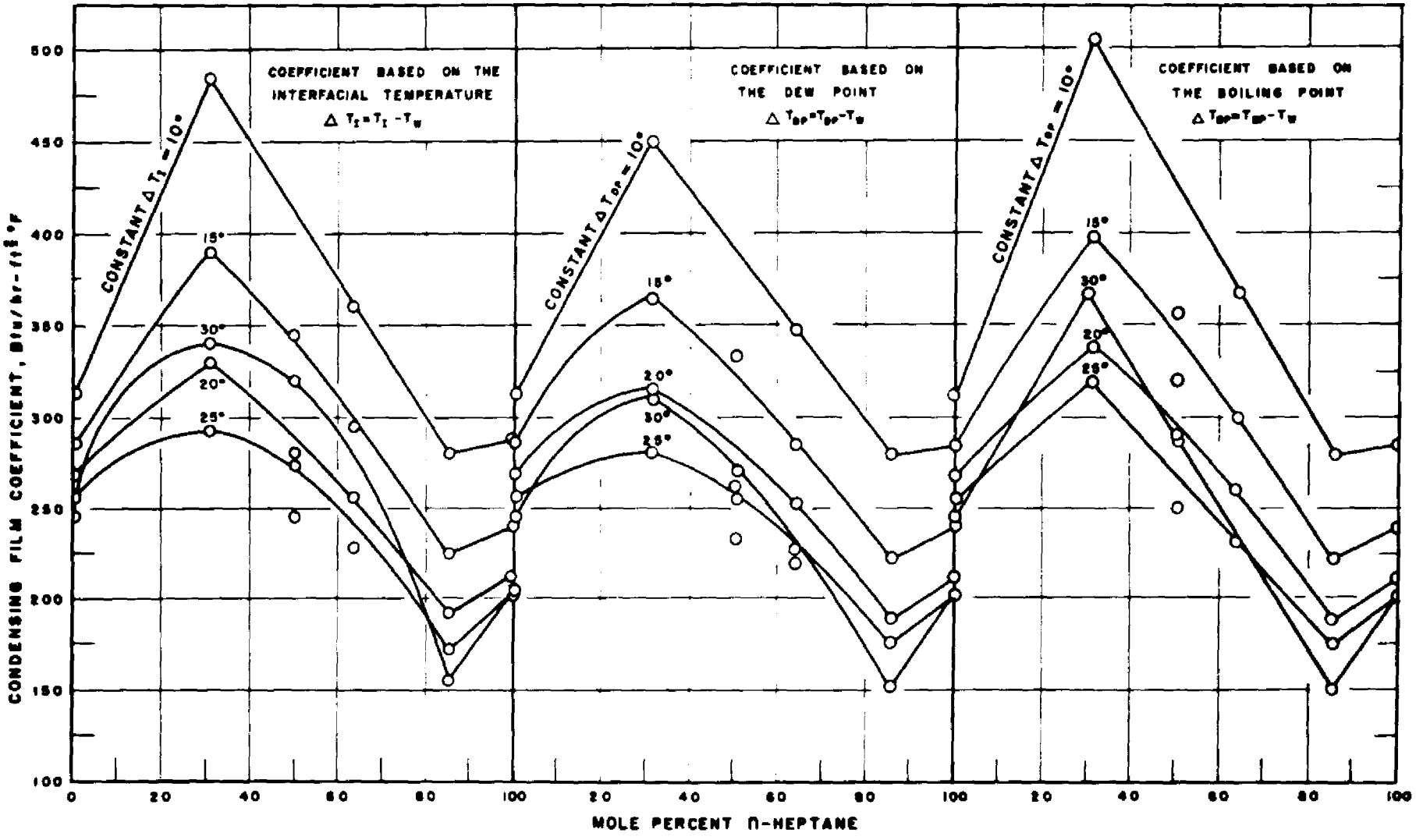


FIGURE 33. VARIATION OF CONDENSING FILM COEFFICIENT WITH COMPOSITION OF BINARY MIXTURES AT CONSTANT TEMPERATURE DIFFERENCE
(n-HEPTANE – TOLUENE SYSTEM)

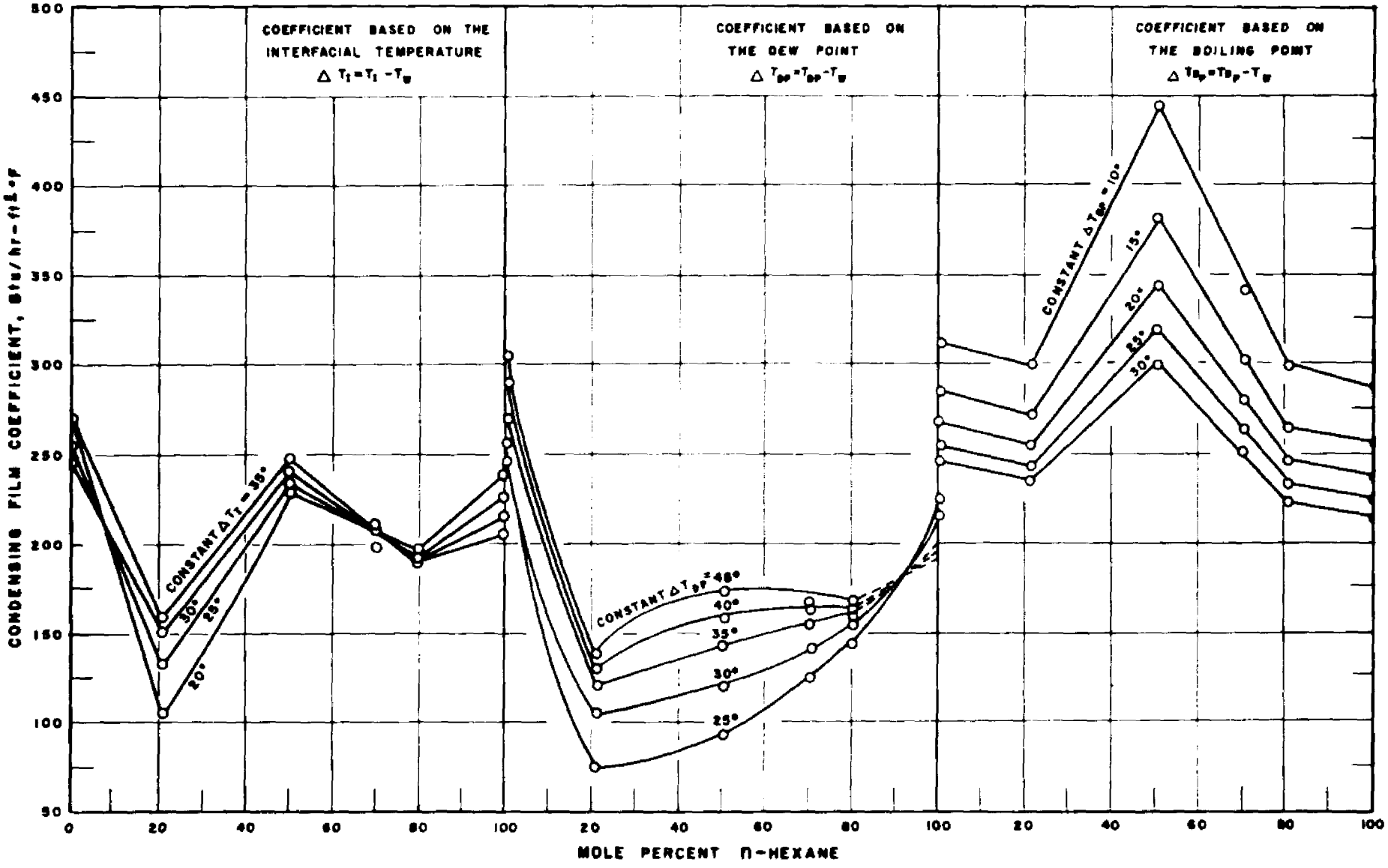


FIGURE 34. VARIATION OF CONDENSING FILM COEFFICIENT WITH COMPOSITION OF BINARY MIXTURES AT CONSTANT TEMPERATURE DIFFERENCE
(n-HEXANE - TOLUENE SYSTEM)

mole per cent n-heptane is the reverse true.

Quite different from the two previously discussed systems are the curves for n-hexane-toluene system, Figure 34. For curves based on dew point and interfacial temperatures, the temperature difference parameters are reverse of the other two systems. In both benzene-n-heptane and n-heptane-toluene systems the magnitude of the parameter decreases with increase in h_I and h_{DP} at constant composition, whereas, the case is reverse for n-hexane-toluene system. This different behavior between n-hexane-toluene mixtures and mixtures of benzene-n-heptane and n-heptane-toluene is also shown in plot of heat transfer coefficient versus temperature drop across the film, for example, Figures 18, 23, and 27. For mixtures of benzene-n-heptane and n-heptane-toluene, the heat transfer coefficient decreases as the temperature drop across the film increases as do the pure components. However, for mixtures of n-hexane-toluene, the heat transfer coefficient based on the dew point and the interface temperature increases as the temperature drop across the film increases. The curves based on dew point temperature show a dip at 21.1 mole per cent. At a concentration of 21.1 to 80.0 mole per cent the curves for 45° and 40° temperature drop show an upward curvature, a straight line for 35°, and a hyperbolic curve for 30° and 25° temperature drop. Again, just as for the other two binary systems, reason for the dips which are present for n-hexane-toluene system is that at concentration that dips occur the heat transfer coefficients are smaller than those of the pure components. The curves based on boiling point temperature show a sharp peak at 50.2 mole per cent n-hexane. Referring to Figure 28, it is

noticed at this concentration the heat transfer data based on boiling point temperature are higher than those of the pure components. This can be generalized as whenever the heat transfer coefficients, based on one of the temperature differences, are smaller than those of the pure components, there is a dip in the heat transfer coefficient-composition curves at constant temperature drop across the film and there exist a peak in these curves if the heat transfer coefficients are greater than those of the pure components.

The sensible heat transferred by vapor, cooling from its temperature to the interface temperature, and the latent heat of condensation evolved at the interface were calculated. These are summarized in Table XI-A, B, and C. In solving Equation 20 in Chapter II, to determine the sensible heat transferred, it was required to know the gas-film heat transfer coefficient. Equation 25 and Figure 46 were used to determine the gas-film heat transfer coefficient. The magnitude of the coefficient for the three binary systems was small, less than 0.35 Btu/Hr.Ft.^{2} °F. Its variation with composition of vapor at constant Reynolds number, for the binary mixtures studied, is less than 5 per cent; therefore, it is considered negligible.

It is noticed from Table XI that the magnitude of term C_o is large, therefore the ratio of $C_o/1-e^{-C_o}$ used in Equation 20 to determine the gas-film heat transfer coefficient is essentially equal to C_o. This corresponds to the fact that the major portion of the condensate was carried to the interface by bulk flow rather than by diffusion. The same conclusion follows for heat transfer: the sensible heat transferred

TABLE XI

RATE OF SENSIBLE HEAT, LATENT HEAT OF CONDENSATION,
AND TOTAL HEAT TRANSFER IN CONDENSING MIXED VAPORS

A BENZENE - n-HEPTANE SYSTEM

Run No.	Rate of Condensation, Gms./Hr.	Rate of Condensation, Lbs./Hr. Ft.²	Reynolds Number	Heat Transfer Factor	Specific Heat		Gas Film Coefficient of Heat Transfer		Temperature °F.			Latent Heat of Condensation at the Interface	Rate of Heat Transfer Btu/Hr. Ft.²		
													Sensible Heat	Latent Heat	Total
	W	G	$D_e G'/\mu_m$	J_H	C_p	$(C_p\mu/k)^{0.56}$	h_g	C_o	T_V	T_I	ΔT	λ_I	q_s	q_{λ_I}	q
					Btu/Lb. °F.		Btu/Hr. Ft.² °F.					Btu/Lb.			
	A 82.57 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER														
1	276.7	15.360	750	0.0085	0.3680	0.8989	0.0817	43.86	179.8	178.3	1.5	166.0	6	2,584	2,590
2	388.9	21.580	1,050	0.0175	0.3680	0.8989	0.1460	51.39	179.8	178.3	1.5	166.0	10	3,578	3,588
3	496.5	27.438	1,360	0.0155	0.3680	0.8989	0.1685	58.00	179.9	178.4	1.5	166.0	14	4,588	4,602
4	601.8	33.368	1,630	0.0143	0.3680	0.8989	0.1857	62.86	179.9	178.3	1.6	166.0	15	5,538	5,553
5	639.7	35.463	1,730	0.0140	0.3680	0.8989	0.1922	64.21	179.9	178.3	1.6	166.0	20	5,987	5,997
6	762.5	42.272	2,060	0.0129	0.3680	0.8989	0.2111	69.68	179.8	178.3	1.5	166.0	22	7,048	7,070
7	261.5	19.083	930	0.0185	0.3680	0.8989	0.1344	49.23	179.9	178.4	1.5	166.0	10	3,160	3,170
8	446.5	88.753	1,200	0.0168	0.3680	0.8989	0.1571	54.83	179.8	178.4	1.4	166.0	12	4,128	4,140
9	586.0	30.869	1,670	0.0149	0.3680	0.8989	0.1766	60.33	179.9	178.4	1.5	166.0	16	5,288	5,304
10	682.5	44.706	2,270	0.0124	0.3680	0.8989	0.2242	72.69	179.8	178.4	1.4	166.0	26	7,751	7,777
	B 70.37 MOLE PER CENT BENZENE FOR y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER														
1	248.6	13.782	680	0.0214	0.3687	0.9047	0.1208	42.27	184.8	180.9	3.9	163.0	10	2,246	2,256
2	232.7	12.900	640	0.0220	0.3687	0.9047	0.1156	41.14	184.0	180.8	3.2	163.0	9	2,108	2,117
3	268.5	14.880	860	0.0195	0.3687	0.9047	0.1327	46.90	184.3	180.8	3.5	163.0	11	2,751	2,762
4	496.5	27.438	1,360	0.0155	0.3687	0.9047	0.1731	58.39	184.6	180.9	3.7	163.0	17	4,068	4,085
5	444.6	24.648	1,220	0.0162	0.3687	0.9047	0.1667	55.85	184.1	180.8	3.3	163.0	15	3,917	3,932
6	576.8	31.976	1,590	0.0134	0.3687	0.9047	0.1746	67.52	184.8	180.9	3.9	163.0	20	5,112	5,132
7	625.8	34.718	1,720	0.0139	0.3687	0.9047	0.1868	65.10	184.2	180.8	3.4	163.0	20	4,884	4,904
8	625.4	34.337	1,690	0.0141	0.3687	0.9047	0.1861	64.16	184.2	180.7	3.5	163.0	20	4,836	4,856
9	711.0	39.416	1,960	0.0132	0.3687	0.9047	0.2121	68.52	184.1	180.9	3.2	163.0	23	5,355	5,378
10	708.3	39.266	1,950	0.0132	0.3687	0.9047	0.2103	68.84	184.7	180.9	3.8	163.0	23	5,393	5,416
11	663.2	35.678	1,770	0.0139	0.3687	0.9047	0.1964	66.90	184.1	181.0	3.1	163.0	21	5,018	5,039
12	386.6	21.371	1,060	0.0175	0.3687	0.9047	0.1524	51.79	184.7	181.0	3.7	163.0	12	3,389	3,401
	C 54.5% MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER														
1	285.0	13.582	690	0.0212	0.3976	0.9093	0.1259	42.69	190.3	184.8	5.5	158.4	10	2,351	2,361
2	336.5	18.654	960	0.0188	0.3976	0.9093	0.1500	49.44	190.4	184.9	5.5	158.4	12	2,935	2,947
3	441.3	24.464	1,260	0.0160	0.3976	0.9093	0.1711	56.85	190.6	185.1	5.5	158.4	15	3,855	3,870
4	552.5	30.669	1,550	0.0146	0.3976	0.9093	0.1955	62.89	191.2	184.9	6.3	158.4	19	4,978	4,997
5	578.3	32.060	1,620	0.0142	0.3976	0.9093	0.1990	64.05	190.6	184.9	5.7	158.4	19	5,159	5,178
6	627.0	35.314	1,790	0.0137	0.3976	0.9093	0.2115	66.38	191.0	185.0	6.0	158.4	21	5,574	5,595
7	718.5	39.832	2,020	0.0130	0.3976	0.9093	0.2304	69.95	190.7	185.0	5.7	158.4	24	6,218	6,242
8	732.9	40.630	2,060	0.0128	0.3976	0.9093	0.2276	71.04	190.1	185.0	7.1	158.4	24	6,358	6,382
9	330.2	18.305	930	0.0185	0.3976	0.9093	0.1480	49.17	190.3	185.0	5.3	158.4	12	2,958	2,970
10	580.2	32.165	1,630	0.0142	0.3976	0.9093	0.1997	64.04	190.7	185.0	5.7	158.4	17	5,395	5,412
	D 32.98 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER														
1	357.5	14.275	760	0.0205	0.4366	0.9172	0.1386	44.76	198.8	195.3	3.5	148.8	11	2,184	2,195
2	379.2	21.082	1,100	0.0171	0.4366	0.9172	0.1703	53.65	198.9	195.4	3.5	148.8	14	3,127	3,141
3	317.3	17.590	920	0.0186	0.4366	0.9172	0.1550	49.32	198.5	194.8	3.7	148.8	12	2,817	2,829
4	583.0	32.320	1,690	0.0160	0.4366	0.9172	0.2166	65.51	198.6	194.6	4.0	148.8	18	4,888	4,906
5	653.3	36.217	1,890	0.0134	0.4366	0.9172	0.2299	68.16	198.5	194.7	3.8	148.8	20	5,158	5,178
6	753.0	41.744	2,180	0.0126	0.4366	0.9172	0.2658	72.80	198.5	194.6	3.9	148.8	23	6,010	6,033
7	684.8	37.993	1,980	0.0147	0.4366	0.9172	0.2086	62.80	198.7	194.7	4.0	148.8	21	4,880	4,901
8	645.4	35.814	1,870	0.0157	0.4366	0.9172	0.2053	59.63	198.6	194.7	3.9	148.8	20	4,738	4,758
9	695.9	23.411	1,210	0.0168	0.4366	0.9172	0.1818	56.63	198.7	194.7	4.0	148.8	17	4,232	4,249
10	593.3	16.880	850	0.0198	0.4366	0.9172	0.1479	47.78	198.9	194.5	4.4	148.8	13	3,349	3,362

TABLE XI

RATE OF SENSIBLE HEAT, LATENT HEAT OF CONDENSATION, AND TOTAL HEAT TRANSFER IN CONDENSING MIXED VAPORS

B n-HEPTANE - TOLUENE SYSTEM

Run No.	Rate of Condensation: Gm./Hr. W	Rate of Condensation: Lbs./Hr. Ft.² G	Reynolds Number $D_e G'/\mu$	Heat Transfer Factor J_H	Specific Heat Btu/Lb. °F. C_p	$\left(\frac{C_p \mu}{k}\right)^{0.56}$	Gas Film Coefficient of Heat Transfer Btu/Hr. Ft.² °F. h_g	C.	Temperature °F.: T_v	Temperature °F.: T_i	Temperature °F.: ΔT	Latent Heat of Condensation at the Interface Btu/Lb. λ_i	Rate of Heat Transfer Btu/Hr. Ft.²: Sensible Heat q_s	Rate of Heat Transfer Btu/Hr. Ft.²: Latent Heat q_{λ_i}	Rate of Heat Transfer Btu/Hr. Ft.²: Total Q
				A	85.68 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	337.2	17.595	940	0.0184	0.4668	0.9373	0.1609	50.95	210.8	210.3	0.5	161.0	4	2,839	2,843
2	357.1	19.797	1,050	0.0175	0.4662	0.9373	0.1723	53.56	210.9	210.3	0.6	161.0	5	2,791	2,796
3	406.7	22.546	1,200	0.0164	0.4662	0.9373	0.1839	57.15	210.8	210.3	0.5	161.0	5	3,179	3,184
4	403.6	22.361	1,190	0.0165	0.4662	0.9373	0.1833	56.82	210.8	210.3	0.5	161.0	5	3,150	3,155
5	385.0	21.343	1,140	0.0167	0.4662	0.9373	0.1773	56.22	210.7	210.3	0.4	161.0	4	3,009	3,013
6	483.0	26.776	1,430	0.0150	0.4668	0.9373	0.1997	62.50	210.9	210.3	0.6	161.0	7	3,775	3,782
7	530.1	29.387	1,560	0.0145	0.4668	0.9373	0.2118	64.65	210.9	210.3	0.6	161.0	8	4,143	4,151
8	550.5	30.518	1,630	0.0142	0.4662	0.9373	0.2155	66.02	211.2	210.3	0.9	161.0	13	4,288	4,301
9	598.0	32.930	1,750	0.0138	0.4662	0.9373	0.2260	67.93	211.0	210.3	0.7	161.0	11	4,523	4,534
10	605.0	34.468	1,850	0.0135	0.4662	0.9373	0.2326	69.44	210.9	210.3	0.6	161.0	9	4,598	4,607
11	682.5	34.510	1,860	0.0136	0.4662	0.9373	0.2334	68.93	210.8	210.3	0.5	161.0	8	4,595	4,603
12	759.4	42.094	2,260	0.0125	0.4662	0.9373	0.2617	74.98	210.8	210.3	0.5	161.0	10	5,705	5,715
13	733.4	40.658	2,170	0.0126	0.4662	0.9373	0.2548	74.39	210.9	210.3	0.6	161.0	11	5,734	5,745
14	852.6	47.266	2,520	0.0119	0.4662	0.9373	0.2797	78.78	210.8	210.3	0.5	161.0	11	6,328	6,339
				B	64.19 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	421.2	23.350	1,210	0.0163	0.4386	0.9460	0.1784	58.06	213.8	212.8	1.0	165.6	10	3,888	3,898
2	523.0	28.994	1,500	0.0147	0.4386	0.9460	0.1976	64.35	213.8	212.8	1.0	165.6	12	4,803	4,815
3	521.6	28.916	1,500	0.0147	0.4386	0.9460	0.1970	64.37	213.8	212.8	1.0	165.6	12	4,802	4,814
4	652.4	36.178	1,880	0.0134	0.4386	0.9460	0.2253	70.62	213.8	212.8	1.0	165.6	17	5,888	5,905
5	727.5	40.331	2,090	0.0129	0.4386	0.9460	0.2412	73.33	213.8	212.8	1.0	165.6	17	6,057	6,074
6	725.7	40.231	2,090	0.0129	0.4386	0.9460	0.2406	73.33	213.8	212.8	1.0	165.6	17	6,047	6,064
7	861.6	47.765	2,480	0.0120	0.4386	0.9460	0.2657	78.84	213.8	212.8	1.0	165.6	21	6,956	6,977
8	569.4	31.566	1,640	0.0142	0.4386	0.9460	0.2078	66.62	213.8	212.8	1.0	165.6	14	5,396	5,410
9	990.2	54.894	2,850	0.0111	0.4386	0.9460	0.2876	83.72	213.8	212.8	1.0	165.6	24	7,998	8,022
10	1113.0	61.702	3,200	0.0108	0.4386	0.9460	0.3089	87.61	213.8	212.8	1.0	165.6	27	8,983	9,010
				C	50.32 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	338.1	18.189	920	0.0185	0.4212	0.9488	0.1496	51.28	216.8	215.8	1.0	169.5	7	2,519	2,526
2	425.8	23.605	1,190	0.0164	0.4212	0.9488	0.1718	57.87	217.0	215.9	1.1	169.5	11	3,329	3,340
3	382.0	21.122	1,070	0.0172	0.4212	0.9488	0.1613	55.15	217.1	215.8	1.3	169.5	11	3,158	3,169
4	518.0	28.716	1,450	0.0150	0.4212	0.9488	0.1912	63.26	216.8	215.9	0.9	169.5	10	4,078	4,088
5	488.5	27.081	1,370	0.0153	0.4212	0.9488	0.1839	62.02	216.8	215.9	0.9	169.5	10	3,912	3,922
6	645.7	36.905	1,870	0.0135	0.4212	0.9488	0.2211	70.30	217.0	216.0	1.0	169.5	15	5,512	5,527
7	598.8	33.196	1,680	0.0140	0.4212	0.9488	0.2063	67.77	217.0	215.9	1.1	169.5	15	5,145	5,160
8	792.8	43.951	2,220	0.0125	0.4212	0.9488	0.2439	75.90	217.0	215.9	1.1	169.5	18	6,578	6,596
9	800.0	44.350	2,260	0.0125	0.4212	0.9488	0.2461	75.90	216.9	215.9	1.0	169.5	18	6,648	6,666
10	1084.0	60.094	3,060	0.0110	0.4212	0.9488	0.2934	86.27	217.0	215.9	1.1	169.5	28	8,984	9,012
				D	31.19 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	477.2	26.455	1,290	0.0159	0.3964	0.9553	0.1765	60.00	222.3	220.2	2.1	153.4	11	4,568	4,579
2	532.7	29.532	1,440	0.0150	0.3964	0.9553	0.1898	63.69	222.6	220.2	2.4	153.4	14	5,530	5,544
3	578.0	32.043	1,560	0.0145	0.3964	0.9553	0.1987	66.91	222.3	220.3	2.0	153.4	13	5,268	5,281
4	689.7	38.235	1,860	0.0135	0.3964	0.9553	0.2268	70.75	222.8	220.4	2.4	153.4	18	6,988	7,006
5	689.7	38.235	1,860	0.0135	0.3964	0.9553	0.2268	70.75	222.5	220.3	2.2	153.4	17	6,988	7,005
6	788.8	43.396	2,110	0.0128	0.3964	0.9553	0.2505	74.48	222.1	220.3	1.8	153.4	19	7,691	7,710
7	856.6	47.484	2,310	0.0122	0.3964	0.9553	0.2698	77.11	222.6	220.3	2.3	153.4	23	8,119	8,142
8	1130.0	62.753	3,100	0.0110	0.3964	0.9553	0.3910	86.86	222.6	220.3	2.3	153.4	28	9,788	9,816
9	1553.4	86.086	4,190	0.0098	0.3964	0.9553	0.3497	97.49	222.4	220.3	2.1	153.4	71	13,159	13,230
10	1650.0	90.385	3,910	0.0100	0.3964	0.9533	0.3335	95.55	222.2	220.3	1.9	153.4	60	12,151	12,211

TABLE XI

RATE OF SENSIBLE HEAT, LATENT HEAT OF CONDENSATION, AND TOTAL HEAT TRANSFER IN CONDENSING MIXED VAPORS

C n-HEXANE - TOLUENE SYSTEM

Run No.	Rate of Condensation: Gms./Min. W	Rate of Condensation: Lbs./Hr. Ft.² G	Reynolds Number $D_e G'/\mu_m$	Heat Transfer Factor J_H	Specific Heat: Btu/Lb. °F. C_p	$\left(\frac{C_p \mu}{k}\right)^{0.56}$	Gas Film Coefficient of Heat Transfer: Btu/Hr. Ft.² °F. h_g	C_o	Temperature °F.: T_V	Temperature °F.: T_I	Temperature °F.: ΔT	Latent Heat of Condensation at the Interface: Btu/Lb. λ_I	Rate of Heat Transfer Btu/Hr. Ft.²: Sensible Heat q_s	Rate of Heat Transfer Btu/Hr. Ft.²: Latent Heat q_{λ_I}	Rate of Heat Transfer Btu/Hr. Ft.²: Total q
					A 80.40 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	848.9	47.062	2,530	0.0118	0.4370	0.9110	0.2648	77.66	176.7	168.0	8.7	159.0	179	7,482	7,661
2	865.0	47.954	2,580	0.0117	0.4370	0.9110	0.2691	77.87	175.8	168.5	7.3	159.0	153	7,625	7,777
3	682.6	37.842	2,040	0.0130	0.4370	0.9110	0.2360	70.07	176.1	168.4	7.7	159.0	127	6,017	6,144
4	616.5	34.177	1,840	0.0135	0.4370	0.9110	0.2213	67.49	176.6	169.0	7.6	159.0	113	5,434	5,547
5	563.7	31.250	1,680	0.0141	0.4370	0.9110	0.2113	64.63	175.8	169.3	6.5	159.0	89	4,969	5,058
6	511.1	29.476	1,590	0.0145	0.4370	0.9110	0.2050	62.83	175.2	169.2	6.0	159.0	77	4,686	4,763
7	501.8	27.819	1,500	0.0148	0.4370	0.9110	0.1975	61.55	176.0	169.2	6.8	159.0	82	4,423	4,505
8	360.0	18.868	1,010	0.0177	0.4370	0.9110	0.1600	51.18	175.4	168.0	7.4	159.0	61	2,996	3,057
9	299.6	16.609	890	0.0190	0.4370	0.9110	0.1513	47.97	175.8	168.0	7.8	159.0	56	2,840	2,896
10	388.5	21.537	1,160	0.0165	0.4370	0.9110	0.1704	55.23	175.3	168.3	7.0	159.0	66	3,436	3,490
					B 70.93 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	804.7	44.610	2,360	0.0128	0.4281	0.9154	0.2587	73.82	185.2	175.2	10.0	162.0	191	7,227	7,418
2	756.7	41.950	2,200	0.0126	0.4281	0.9154	0.2472	72.54	184.9	175.2	9.7	162.0	174	6,796	6,970
3	680.0	37.697	1,960	0.0131	0.4281	0.9154	0.2309	69.90	184.8	176.0	8.8	162.0	142	6,107	6,249
4	612.8	33.972	1,780	0.0137	0.4281	0.9154	0.2176	66.83	185.6	174.3	11.3	162.0	184	5,503	5,687
5	526.1	29.166	1,530	0.0146	0.4281	0.9154	0.1991	62.71	184.7	175.0	9.7	162.0	121	4,725	4,846
6	421.1	23.345	1,220	0.0162	0.4281	0.9154	0.1768	56.52	185.0	175.2	9.8	162.0	98	3,782	3,880
7	282.4	15.655	820	0.0195	0.4281	0.9154	0.1427	46.96	185.3	175.2	10.1	162.0	67	2,536	2,603
8	251.0	13.915	730	0.0206	0.4281	0.9154	0.1360	44.45	184.7	175.0	9.7	162.0	58	2,254	2,312
9	484.9	25.773	1,350	0.0155	0.4281	0.9154	0.1868	59.06	185.9	174.5	11.3	162.0	124	4,176	4,300
					C 51.30 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	812.0	45.015	2,270	0.0124	0.4088	0.9232	0.2671	74.47	201.2	186.6	14.6	165.5	268	7,450	7,718
2	735.9	40.796	2,060	0.0129	0.4088	0.9232	0.2330	71.57	201.3	187.0	14.3	165.5	238	6,752	6,990
3	692.0	38.363	1,930	0.0132	0.4088	0.9232	0.2282	69.95	201.3	186.8	14.5	165.5	227	6,349	6,576
4	604.8	33.529	1,690	0.0140	0.4088	0.9232	0.2078	65.96	201.7	187.5	14.2	165.5	195	5,599	5,794
5	548.1	30.385	1,530	0.0146	0.4088	0.9232	0.1948	63.24	201.9	187.0	14.9	165.5	185	5,088	5,273
6	500.7	27.758	1,400	0.0152	0.4088	0.9232	0.1868	60.76	202.4	186.2	15.2	165.5	172	4,594	4,766
7	467.9	25.939	1,300	0.0157	0.4088	0.9232	0.1803	58.81	200.8	186.6	14.2	165.5	150	4,293	4,443
8	355.8	19.725	990	0.0180	0.4088	0.9232	0.1572	51.30	201.6	185.4	16.2	165.5	130	3,288	3,394
9	337.7	18.721	940	0.0185	0.4088	0.9232	0.1533	49.99	201.7	185.8	15.9	165.5	121	3,099	3,220
10	301.8	16.731	840	0.0194	0.4088	0.9232	0.1437	47.60	200.2	185.8	14.4	165.5	98	2,749	2,847
					D 21.92 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER										
1	683.6	37.897	1,800	0.0136	0.3784	0.9315	0.2094	68.48	221.4	212.3	9.1	165.0	130	6,253	6,383
2	594.0	32.930	1,560	0.0145	0.3784	0.9315	0.1939	64.26	221.7	212.3	9.4	165.0	117	5,433	5,550
3	530.8	29.426	1,400	0.0152	0.3784	0.9315	0.1817	61.28	220.6	212.9	7.7	165.0	86	4,855	4,941
4	492.5	27.303	1,300	0.0157	0.3784	0.9315	0.1741	59.34	220.9	212.6	8.3	165.0	86	4,504	4,590
5	417.4	23.140	1,100	0.0170	0.3784	0.9315	0.1598	54.79	221.3	212.7	8.6	165.0	75	3,818	3,893
6	344.9	19.120	910	0.0186	0.3784	0.9315	0.1444	50.10	221.1	212.3	8.8	165.0	64	3,154	3,218
7	193.8	10.743	510	0.0245	0.3784	0.9315	0.1069	38.03	221.8	212.6	9.2	165.0	37	1,773	1,810
8	152.3	8.443	400	0.0275	0.3784	0.9315	0.0943	33.88	220.9	213.2	7.7	165.0	25	1,343	1,418
9	266.7	14.674	700	0.0211	0.3784	0.9315	0.1258	44.14	221.3	213.4	7.9	165.0	44	2,421	2,465

from the vapor by convection, for the three binary systems studied was less than 3 per cent of the total heat transferred.

MASS TRANSFER DATA

The second objective of this work was to study mass transfer operation in condensing binary vapors. In particular, it was desired to determine individual and over-all gas-phase mass transfer coefficients, numbers and heights of the individual and over-all gas-phase transfer units. The objective was also to confirm and determine the extent of the liquid-film resistance, which interpretation of heat transfer data indicates to be present.

The data based on the individual gas-phase resistance are summarized in Table XII-A, B, and C and those based on the over-all gas-phase resistance are shown in Table XIII-A, B, and C.

The individual gas-phase mass transfer coefficients were calculated by the use of Equation 53. The values of k_{g1} which were determined independently from Equation 26 are also tabulated in Table XII, for comparison. For the three binary systems, k_g differs from 0.4 to 40 per cent from k_{g1}. The average difference for each binary system is, about 10 per cent. It is believed this is a satisfactory check on the data.

The number of individual gas-phase transfer units for the three binary systems were found to vary between 2 to 4, depending on the binary system and composition. The height of a gas-phase transfer unit varied between 0.15 to 0.35 feet, again varying with the system and composition.

The working equation developed in Appendix C, Section II-C-5,

TABLE XII

NUMBER OF INDIVIDUAL GAS-PHASE TRANSFER UNITS, HEIGHT OF AN INDIVIDUAL GAS-PHASE TRANSFER UNIT, AND GAS-PHASE MASS TRANSFER COEFFICIENT

A BENZENE - n-HEPTANE SYSTEM

Run No.	Vapor Flow Rate, Lb.-Moles/Hr. V_2	Mole Per Cent Gas-Phase In Vapor y_{A1}	y_{A3}	y_{A2}	No. of Individual Gas-Phase Transfer Units N_g	Height of An Individual Gas-Phase Transfer Unit, Ft. H_g	Gas-Phase Mass Transfer Coefficient, Lb.-Moles/Hr. Ft.^{2} Mole Fraction k_g	k_{g1}	Per Cent Difference*
		A 89.57 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0211	85.33	82.72	85.2	2.996	0.217	0.0265	0.0266	- 0.4
2	0.0210	85.28	82.69	85.2	3.473	0.187	0.0306	0.0344	- 2.5
3	0.0210	85.27	82.67	85.2	3.656	0.178	0.0318	0.0359	- 11.4
4	0.0232	85.37	82.47	85.3	3.719	0.175	0.0362	0.0402	- 9.9
5	0.0192	85.25	82.46	85.2	4.017	0.162	0.0323	0.0410	- 21.2
6	0.0178	85.34	82.54	85.3	4.244	0.153	0.0317	0.0455	- 30.3
7	0.0214	84.88	82.59	84.8	3.350	0.194	0.0300	0.0298	+ 0.7
8	0.0219	84.86	82.67	84.8	3.593	0.181	0.0330	0.0338	- 2.3
9	0.0219	84.86	82.47	84.8	3.657	0.178	0.0338	0.0381	- 11.3
10	0.0187	85.24	82.44	85.2	4.244	0.153	0.0333	0.0487	- 31.6
		B 70.37 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0190	75.96	70.26	75.7	3.084	0.211	0.0241	0249	- 3.2
2	0.0200	76.30	70.90	76.0	2.958	0.220	0.0248	0242	+ 2.5
3	0.0214	76.26	70.21	76.0	3.144	0.207	0.0298	0278	+ 5.0
4	0.0236	75.96	70.37	75.8	3.549	0.183	0.0351	0356	- 1.4
5	0.0240	76.40	70.39	76.2	3.399	0.191	0.0342	0338	+ 1.1
6	0.0242	75.95	70.35	75.8	3.616	0.180	0.0365	0389	- 6.2
7	0.0250	76.35	70.35	76.2	3.685	0.176	0.0386	0406	- 4.9
8	0.0253	76.55	70.46	76.4	3.699	0.175	0.0393	0401	- 2.0
9	0.0205	75.84	70.34	75.7	3.666	0.177	0.0438	0439	- 0.2
10	0.0205	75.83	70.44	75.7	3.719	0.175	0.0445	0438	+ 3.0
11	0.0253	75.70	70.48	75.6	3.950	0.164	0.0419	0455	- 7.9
12	0.0239	75.80	70.31	75.6	3.308	0.196	0.0332	0413	+ 6.1
		C 54.9% MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0276	64.64	55.44	64.0	2.694	0.241	0.0308	0249	+ 23.7
2	0.0264	64.46	55.24	64.0	2.994	0.217	0.0331	0291	+ 13.8
3	0.0276	63.75	54.85	63.4	3.232	0.201	0.0334	0355	+ 5.3
4	0.0238	64.46	55.15	64.2	3.574	0.182	0.0357	0380	- 6.0
5	0.0238	64.25	54.83	64.0	3.685	0.179	0.0362	0384	- 5.7
6	0.0261	64.15	54.65	63.9	3.633	0.179	0.0398	0408	- 2.5
7	0.0307	63.96	54.65	63.7	3.574	0.182	0.0460	0437	+ 5.3
8	0.0269	64.33	54.73	64.1	3.727	0.174	0.0420	0440	- 4.5
9	0.0278	64.30	54.99	63.8	3.151	0.206	0.0340	0289	+ 17.6
10	0.0246	64.25	54.85	64.0	3.623	0.179	0.0374	0385	- 2.9
		D 32.98 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0292	38.93	33.06	38.5	2.611	0.249	0.0319	0254	+ 13.8
2	0.0262	38.58	32.56	38.3	3.065	0.212	0.0337	0307	+ 9.8
3	0.0318	40.25	33.39	39.8	2.718	0.239	0.0354	0283	+ 25.0
4	0.0292	40.44	33.80	40.2	3.404	0.191	0.0417	0385	+ 8.3
5	0.0238	40.17	33.13	40.0	3.719	0.175	0.0371	0409	- 9.2
6	0.0314	40.42	32.83	40.2	3.573	0.182	0.0446	0428	+ 4.0
7	0.0286	40.28	32.58	40.0	3.310	0.196	0.0397	0363	+ 9.4
8	0.0289	40.12	32.72	39.8	3.137	0.207	0.0380	0337	+ 12.7
9	0.0263	40.20	32.92	39.9	3.185	0.204	0.0365	0325	+ 12.3
10	0.318	40.71	32.45	40.2	2.792	0.233	0.0374	0271	+ 35.6
		Average per cent difference							9.3

* Per cent difference is defined as $\frac{k_g - k_{g1}}{k_{g1}}$ (100)

TABLE XII

NUMBER OF INDIVIDUAL GAS-PHASE TRANSFER UNITS, HEIGHT OF AN INDIVIDUAL GAS-PHASE TRANSFER UNIT, AND GAS-PHASE MASS TRANSFER COEFFICIENT

B n-HEPTANE - TOLUENE SYSTEM

Run No.	Vapor Flow Rate, Lb. - Moles / Hr. V_2	Mole Per Cent n-Heptane In Vapor y_{A1}	y_{A1}	y_{A2}	No. of Individual Gas-Phase Transfer Units N_g	Height of An Individual Gas-Phase Transfer Unit, Ft. H_g	Gas-Phase Mass Transfer Coefficient, Lb.-Moles / Hr., Ft.², Mole Fraction k_g	k_{g1}	Per Cent * Difference
		A 85.68 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0480	86.81	85.70	86.7	2.309	0.282	0.0265	0.0260	+ 1.9
2	0.0454	86.68	85.70	86.6	2.503	0.259	0.0276	0.0280	- 1.4
3	0.0507	86.79	85.62	86.7	2.562	0.253	0.0286	0.0299	+ 15.7
4	0.0534	86.80	85.53	86.7	2.539	0.256	0.0366	0.0300	+ 22.0
5	0.0480	86.77	85.78	86.7	2.643	0.246	0.0348	0.0290	+ 20.0
6	0.0454	86.76	85.70	86.7	2.868	0.226	0.0355	0.0325	+ 9.2
7	0.0440	86.76	85.66	86.7	2.905	0.224	0.0336	0.0340	- 1.2
8	0.0480	86.76	85.66	86.7	2.905	0.224	0.0385	0.0349	+ 10.3
9	0.0415	86.65	85.64	86.6	3.002	0.216	0.0322	0.0364	- 11.5
10	0.0454	86.65	85.70	86.6	2.941	0.221	0.0360	0.0375	- 4.0
11	0.0496	86.65	85.75	86.6	2.887	0.225	0.0358	0.0372	- 3.7
12	0.0507	86.64	85.72	86.6	3.132	0.207	0.0466	0.0419	+ 11.2
13	0.0574	86.76	85.66	86.7	2.905	0.224	0.0499	0.0415	+ 20.2
14	0.0600	86.75	85.66	86.7	3.078	0.211	0.0376	0.0449	- 16.2
		B 64.13 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0488	67.98	64.40	67.7	2.545	0.255	0.0301	0.0306	- 1.6
2	0.0433	68.01	64.32	67.8	2.863	0.227	0.0300	0.0343	- 14.3
3	0.0515	67.94	64.34	67.7	2.705	0.240	0.0384	0.0343	+ 12.0
4	0.0516	68.01	64.12	67.8	2.915	0.223	0.0431	0.0386	+ 11.6
5	0.0518	67.89	64.10	67.7	2.990	0.217	0.0450	0.0411	+ 9.5
6	0.0515	67.88	64.13	67.7	3.033	0.214	0.0455	0.0411	+ 10.7
7	0.0558	67.98	63.89	67.8	3.120	0.208	0.0430	0.0450	- 4.5
8	0.0528	68.05	64.06	67.8	2.767	0.235	0.0313	0.0355	- 11.8
9	0.0555	67.96	63.97	67.8	3.213	0.202	0.0448	0.0487	- 8.0
10	0.0583	67.95	63.96	67.8	3.277	0.196	0.0503	0.0522	- 3.6
		C 50.32 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0516	53.49	50.63	53.2	2.286	0.284	0.0265	0.0272	- 2.5
2	0.0487	53.44	50.26	53.2	2.581	0.252	0.0307	0.0309	- 0.6
3	0.0502	53.48	50.28	53.2	2.433	0.267	0.0302	0.0292	+ 3.4
4	0.0524	53.28	50.58	53.1	2.705	0.240	0.0300	0.0342	- 12.3
5	0.0508	53.29	50.58	53.1	2.654	0.245	0.0303	0.0334	- 9.3
6	0.0521	52.95	50.15	52.8	2.923	0.222	0.0334	0.0390	- 14.3
7	0.0521	53.07	50.17	52.9	2.833	0.229	0.0310	0.0369	- 16.0
8	0.0535	53.14	50.14	53.0	3.061	0.212	0.0387	0.0437	- 11.4
9	0.0562	53.23	50.43	53.1	3.066	0.212	0.0403	0.0427	- 5.6
10	0.0590	53.32	50.02	53.2	3.310	0.196	0.0900	0.0515	- 2.9
		D 31.19 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0473	36.21	31.53	35.9	2.721	0.240	0.0308	0.0330	- 6.7
2	0.0564	36.12	31.52	35.8	2.662	0.244	0.0430	0.0348	+ 23.6
3	0.0534	35.99	31.33	35.7	2.773	0.234	0.0421	0.0362	+ 16.3
4	0.0559	36.28	31.08	36.0	2.918	0.223	0.0484	0.0401	+ 20.7
5	0.0529	36.03	31.43	35.8	2.992	0.217	0.0464	0.0400	+ 16.0
6	0.0543	36.01	31.53	35.8	3.056	0.212	0.0400	0.0434	- 8.5
7	0.0571	36.06	30.26	35.8	3.101	0.209	0.0543	0.0451	+ 20.2
8	0.0598	35.96	31.55	35.8	3.312	0.196	0.0431	0.0540	- 20.2
9	0.0640	35.95	30.66	35.8	3.559	0.182	0.0635	0.0643	+ 1.2
10	0.0543	35.93	31.02	35.8	3.627	0.179	0.0526	0.0612	- 14.1
		Average per cent difference							10.5

* Per cent difference is defined as $\frac{k_g - k_{g1}}{k_{g1}}$ (100)

TABLE XII

NUMBER OF INDIVIDUAL GAS-PHASE TRANSFER UNITS, HEIGHT OF AN INDIVIDUAL GAS-PHASE TRANSFER UNIT, AND GAS-PHASE MASS TRANSFER COEFFICIENT

C n-HEXANE - TOLUENE SYSTEM

Run No.	Vapor Flow Rate, Lb. - Moles / Hr. V_2	Mole Per Cent n-Hexane In Vapor y_{A1}	y_{A1}	y_{A2}	No. of Individual Gas-Phase Transfer Units N_g	Height of An Individual Gas-Phase Transfer Unit, Ft. H_g	Gas-Phase Mass Transfer Coefficient, Lb.-Moles / Hr. Ft.^{2} Mole Fraction k_g	k_{g1}	Per Cent * Difference
		A 80.40 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0656	86.92	80.23	86.6	3.036	0.214	0.0436	0.0488	- 10.6
2	0.0596	86.24	80.64	86.0	3.146	0.206	0.0486	0.0490	- 0.8
3	0.0587	86.54	80.15	86.2	2.930	0.222	0.0422	0.0432	- 2.3
4	0.0444	85.97	79.96	85.7	3.099	0.210	0.0377	0.0405	- 6.9
5	0.0567	85.50	80.61	85.2	2.788	0.233	0.0363	0.0388	- 6.4
6	0.0486	85.16	80.56	84.9	2.870	0.226	0.0385	0.0369	+ 4.3
7	0.0570	85.11	80.58	84.8	2.679	0.242	0.0340	0.0366	- 7.1
8	0.0511	86.82	80.27	86.2	2.354	0.276	0.0354	0.0297	+ 19.2
9	0.0630	86.89	80.40	86.1	2.104	0.309	0.0257	0.0285	- 9.9
10	0.0566	86.51	80.62	86.0	2.444	0.266	0.0353	0.0321	- 10.0
		B 70.93 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0774	78.79	70.68	78.3	2.803	0.232	0.0502	0.0477	+ 5.2
2	0.0802	79.15	70.75	78.6	2.723	0.239	0.0516	0.0466	+ 10.7
3	0.0796	78.58	70.58	78.0	2.621	0.248	0.0475	0.0438	+ 8.4
4	0.0894	80.13	70.71	79.3	2.426	0.268	0.0406	0.0415	- 2.1
5	0.0695	79.68	71.22	79.0	2.518	0.258	0.0334	0.0380	- 12.1
6	0.0605	79.43	71.07	78.7	2.435	0.267	0.0337	0.0335	+ 0.6
7	0.0710	79.43	71.55	78.3	1.939	0.335	0.0244	0.0281	- 13.2
8	0.0484	79.57	71.04	78.6	2.172	0.299	0.0282	0.0258	+ 8.5
9	0.0469	80.08	70.77	79.5	2.772	0.234	0.0345	0.0348	- 0.8
		C 51.30 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0615	66.87	50.97	66.1	3.024	0.215	0.0496	0.0478	+ 3.7
2	0.0644	66.34	51.24	65.5	2.885	0.225	0.0496	0.0454	+ 9.3
3	0.0720	66.60	51.22	65.6	2.730	0.238	0.0425	0.0442	- 3.9
4	0.0602	65.94	50.94	65.0	2.766	0.235	0.0452	0.0409	+ 10.5
5	0.0659	66.35	50.96	65.2	2.591	0.251	0.0416	0.0389	+ 7.0
6	0.0615	67.42	51.42	66.2	2.570	0.253	0.0363	0.0378	- 4.0
7	0.0511	66.87	51.27	65.8	2.653	0.245	0.0394	0.0353	+ 11.6
8	0.0494	68.30	51.90	66.9	2.458	0.264	0.0389	0.0309	+ 28.9
9	0.0494	67.66	51.49	66.2	2.402	0.270	0.0398	0.0303	+ 31.3
10	0.0408	67.64	51.55	66.3	2.482	0.262	0.0225	0.0282	- 20.2
		D 21.92 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER							
1	0.0340	32.98	21.28	32.6	3.423	0.190	0.0488	0.0428	+ 14.0
2	0.0527	32.94	21.81	32.3	2.853	0.228	0.0331	0.0396	- 16.4
3	0.0484	32.00	21.87	31.4	2.823	0.230	0.0373	0.0375	- 0.5
4	0.0498	32.52	21.41	31.8	2.733	0.238	0.0371	0.0362	+ 2.5
5	0.0396	32.27	21.26	31.6	2.796	0.232	0.0364	0.0329	+ 10.6
6	0.0360	32.74	21.62	32.0	2.706	0.240	0.0222	0.0299	- 34.7
7	0.0403	33.52	22.72	32.2	2.099	0.310	0.0237	0.0230	+ 3.0
8	0.0410	31.66	22.95	30.3	1.854	0.350	0.0284	0.0204	+ 39.2
9	0.0491	31.20	22.39	30.2	2.173	0.299	0.0248	0.0268	- 7.5
		Average per cent difference							10.5

* Per cent difference is defined as $\frac{k_g - k_{g1}}{k_{g1}}$ (100)

TABLE XIII

NUMBER OF OVER-ALL GAS-PHASE TRANSFER UNITS, HEIGHT OF AN OVER-ALL GAS-PHASE TRANSFER UNIT, AND OVER-ALL MASS TRANSFER COEFFICIENT

A BENZENE - n-HEPTANE SYSTEM

Run No.	Vapor Flow Rate, Lb. - Moles / Hr. V_2	Mole Per Cent Benzene in Vapor y_A^*	y_{A1}	y_{A2}	No. of Over-All Gas-Phase Transfer Units N_{og}	Height of An Over-All Gas-Phase Transfer Unit, Ft. H_{og}	Over-All Mass-Transfer Coefficient, Lb.-Moles / Hr. Fr.^{2} Mole Fraction K_g
		A 82.57 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0211	86.6	82.72	85.2	1.018	0.638	0.0090
2	0.0210	86.6	82.69	85.2	1.026	0.633	0.0090
3	0.0210	86.6	82.67	85.2	1.031	0.630	0.0091
4	0.0232	86.4	82.47	85.3	1.272	0.511	0.0124
5	0.0192	86.4	82.46	85.2	1.187	0.547	0.0095
6	0.0178	86.4	82.54	85.3	1.254	0.518	0.0094
7	0.0214	86.4	82.59	84.8	0.866	0.750	0.0078
8	0.0219	86.6	82.67	84.8	0.780	0.833	0.0071
9	0.0219	86.4	82.47	84.8	0.897	0.724	0.0082
10	0.0187	86.4	82.44	85.2	1.192	0.545	0.0094
		B 70.37 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0190	77.6	70.26	75.7	1.350	0.481	0.0105
2	0.0200	77.7	70.50	76.0	1.441	0.451	0.0121
3	0.0214	77.6	70.21	76.0	1.528	0.425	0.0149
4	0.0236	77.7	70.37	75.8	1.348	0.482	0.0133
5	0.0240	77.7	70.39	76.2	1.582	0.411	0.0159
6	0.0242	77.7	70.35	75.8	1.351	0.481	0.0137
7	0.0250	77.7	70.35	76.2	1.587	0.409	0.0166
8	0.0253	77.8	70.46	76.4	1.655	0.393	0.0175
9	0.0285	77.7	70.34	75.7	1.302	0.499	0.0155
10	0.0285	77.8	70.44	75.7	1.253	0.519	0.0150
11	0.0253	77.8	70.48	75.6	1.200	0.541	0.0127
12	0.0239	77.7	70.31	75.6	1.256	0.517	0.0126
		C 54.94 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0276	67.0	55.44	64.0	1.347	0.482	0.0156
2	0.0264	67.0	55.24	64.0	1.364	0.476	0.0151
3	0.0276	66.8	54.85	63.4	1.255	0.518	0.0145
4	0.0238	67.0	55.15	64.2	1.441	0.451	0.0144
5	0.0238	66.8	54.83	64.0	1.520	0.427	0.0145
6	0.0261	66.6	54.65	63.9	1.485	0.438	0.0162
7	0.0307	66.6	54.65	63.7	1.414	0.460	0.0182
8	0.0269	66.7	54.73	64.1	1.525	0.426	0.0172
9	0.0278	66.7	54.99	63.8	1.394	0.466	0.0162
10	0.0246	66.8	54.85	64.0	1.451	0.447	0.0149
		D 32.98 MOLE PER CENT BENZENE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0292	48.6	33.06	38.5	0.430	1.511	0.0052
2	0.0262	48.2	32.56	38.3	0.456	1.425	0.0050
3	0.0318	48.8	33.39	39.8	0.537	1.210	0.0071
4	0.0292	48.8	33.20	40.2	0.594	1.094	0.0073
5	0.0238	48.8	33.13	40.0	0.576	1.128	0.0057
6	0.0314	48.7	32.83	40.2	0.623	1.043	0.0082
7	0.0286	48.2	32.58	40.0	0.643	1.011	0.0080
8	0.0289	48.5	32.72	39.8	0.594	1.094	0.0072
9	0.0263	48.6	32.92	39.9	0.588	1.105	0.0065
10	0.0318	48.8	32.45	40.2	0.642	1.012	0.0085

TABLE XIII

NUMBER OF OVER-ALL GAS-PHASE TRANSFER UNITS, HEIGHT OF AN OVER-ALL GAS-PHASE TRANSFER UNIT, AND OVER-ALL MASS TRANSFER COEFFICIENT

B n-HEPTANE - TOLUENE SYSTEM

Run No.	Vapor Flow Rate, Lb. - Moles Hr. V_2		Mole Per Cent n-Heptane in Vapor y_A^*	y_{A1}	y_{A2}	No. of Over-All Gas-Phase Transfer Units N_{og}	Height of An Over-All Gas-Phase Transfer Unit, Ft. H_{og}	Over-All Mass-Transfer Coefficient, Lb.-Moles Hr. Fr.* Mole Fraction K_g
		A	85.68 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0480		88.2	85.70	86.7	0.510	1.274	0.0102
2	0.0454		88.2	85.70	86.6	0.445	1.460	0.0085
3	0.0507		88.2	85.62	86.7	0.542	1.199	0.0136
4	0.0534		88.1	85.53	86.7	0.607	1.071	0.0159
5	0.0480		88.3	85.78	86.7	0.454	1.432	0.0111
6	0.0454		88.2	85.70	86.7	0.510	1.274	0.0116
7	0.0440		88.2	85.66	86.7	0.526	1.236	0.0097
8	0.0480		88.2	85.66	86.7	0.526	1.236	0.0106
9	0.0415		88.2	85.64	86.6	0.469	1.386	0.0082
10	0.0454		88.3	85.70	86.6	0.424	1.533	0.0081
11	0.0494		88.3	85.75	86.6	0.405	1.605	0.0084
12	0.0507		88.3	85.72	86.6	0.417	1.559	0.0088
13	0.0594		88.2	85.66	86.7	0.526	1.236	0.0126
14	0.0600		88.2	85.66	86.7	0.526	1.236	0.0132
		B	64.13 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0488		69.2	64.40	67.7	1.162	0.559	0.0237
2	0.0433		69.2	64.32	67.8	1.247	0.521	0.0226
3	0.0515		69.2	64.34	67.7	1.174	0.553	0.0254
4	0.0516		69.1	64.12	67.8	1.341	0.485	0.0290
5	0.0518		69.1	64.10	67.7	1.271	0.511	0.0276
6	0.0515		69.1	64.13	67.7	1.265	0.514	0.0273
7	0.0558		69.1	63.89	67.8	1.386	0.469	0.0324
8	0.0528		69.0	64.06	67.8	1.413	0.460	0.0313
9	0.0555		69.0	63.97	67.8	1.431	0.454	0.0333
10	0.0583		69.0	63.96	67.8	1.433	0.453	0.0350
		C	50.32 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0516		58.0	50.63	53.2	0.428	1.518	0.0092
2	0.0487		57.7	50.26	53.2	0.502	1.295	0.0102
3	0.0502		57.7	50.28	53.2	0.500	1.300	0.0105
4	0.0524		58.0	50.58	53.1	0.415	1.566	0.0086
5	0.0506		58.0	50.58	53.1	0.415	1.560	0.0080
6	0.0521		57.7	50.15	52.8	0.432	1.504	0.0094
7	0.0521		57.7	50.17	52.9	0.450	1.444	0.0098
8	0.0535		57.7	50.14	53.0	0.474	1.371	0.0106
9	0.0562		57.9	50.43	53.1	0.442	1.470	0.0137
10	0.0590		57.7	50.02	53.2	0.533	1.219	0.0132
		D	31.19 MOLE PER CENT n-HEPTANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0473		40.9	31.53	35.9	0.627	1.036	0.0124
2	0.0564		40.9	31.52	35.8	0.608	1.069	0.0144
3	0.0534		40.7	31.33	35.7	0.627	1.036	0.0140
4	0.0559		40.6	31.08	36.0	0.726	0.895	0.0170
5	0.0529		40.9	31.43	35.8	0.618	1.052	0.0137
6	0.0543		40.9	31.53	35.8	0.607	1.071	0.0141
7	0.0571		40.0	30.76	35.8	0.840	0.774	0.0201
8	0.0598		41.0	31.55	35.8	0.596	1.090	0.0149
9	0.0640		40.0	30.66	35.8	0.798	0.814	0.0214
10	0.0543		40.6	31.02	35.8	0.690	0.942	0.0157

TABLE XIII

NUMBER OF OVER-ALL GAS-PHASE TRANSFER UNITS, HEIGHT OF AN OVER-ALL GAS-PHASE TRANSFER UNIT, AND OVER-ALL MASS TRANSFER COEFFICIENT

C n-HEXANE - TOLUENE SYSTEM

Run No.	Vapor Flow Rate, Lb. - Moles / Hr. V_2	Mole Per Cent n-Hexane in Vapor y_A^*	y_{A1}	y_{A2}	No. of Over-All Gas-Phase Transfer Units N_{og}	Height of An Over-All Gas-Phase Transfer Unit, Ft. H_{og}	Over-All Mass-Transfer Coefficient, Lb.-Moles / Hr. Fr.^{2} Mole Fraction K_g
		A 80.40 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0656	92.2	80.23	86.6	0.758	0.857	0.0208
2	0.0596	92.4	80.64	86.0	0.607	1.071	0.0152
3	0.0587	92.2	80.15	86.2	0.694	0.936	0.0171
4	0.0444	92.0	79.96	85.7	0.647	1.004	0.0120
5	0.0567	92.4	80.61	85.2	0.469	1.385	0.0117
6	0.0486	92.4	80.56	84.9	0.455	1.428	0.0093
7	0.0570	92.4	80.58	84.8	0.441	1.474	0.0105
8	0.0541	92.2	80.27	86.2	0.686	0.947	0.0156
9	0.0630	92.0	80.40	86.1	0.675	0.963	0.0178
10	0.0566	92.3	80.62	86.0	0.616	1.055	0.0148
		B 70.93 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0774	87.6	70.68	78.3	0.597	1.089	0.0194
2	0.0802	87.6	70.75	78.6	0.626	1.038	0.0256
3	0.0796	87.5	70.58	78.0	0.576	1.128	0.0192
4	0.0894	87.5	70.71	79.3	0.692	0.939	0.0268
5	0.0695	87.7	71.22	79.0	0.638	1.019	0.0186
6	0.0605	87.6	71.07	78.7	0.618	1.052	0.0136
7	0.0710	87.7	71.55	78.3	0.540	1.204	0.0149
8	0.0484	87.5	71.04	78.6	0.614	1.058	0.0108
9	0.0469	87.6	70.77	79.5	0.730	0.890	0.0143
		C 51.30 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0615	77.0	50.97	66.1	0.869	0.748	0.0224
2	0.0644	77.2	51.24	65.5	0.796	0.816	0.0215
3	0.0720	77.0	51.22	65.6	0.815	0.797	0.0248
4	0.0602	76.0	50.94	65.0	0.822	0.791	0.0208
5	0.0659	77.0	50.96	65.2	0.790	0.823	0.0218
6	0.0615	77.0	51.42	66.2	0.861	0.755	0.0222
7	0.0511	77.0	51.27	65.8	0.831	0.782	0.0178
8	0.0494	77.4	51.90	66.9	0.886	0.733	0.0183
9	0.0494	77.0	51.49	66.2	0.858	0.757	0.0178
10	0.0408	77.0	51.55	66.3	0.865	0.751	0.0148
		D 21.92 MOLE PER CENT n-HEXANE IN y_1, INCOMING VAPOR TO THE PRIMARY CONDENSER					
1	0.0340	51.0	21.28	32.6	0.479	1.357	0.0068
2	0.0527	51.3	21.81	32.3	0.439	1.480	0.0097
3	0.0484	51.3	21.87	31.4	0.391	1.662	0.0079
4	0.0498	50.7	21.41	31.8	0.438	1.484	0.0091
5	0.0396	50.5	21.26	31.6	0.436	1.491	0.0072
6	0.0360	51.0	21.62	32.0	0.435	1.494	0.0066
7	0.0403	51.5	22.72	32.2	0.399	1.629	0.0067
8	0.0410	51.8	22.95	30.3	0.293	2.218	0.0050
9	0.0491	51.5	22.39	30.2	0.311	2.090	0.0064

Equation 29-C, for calculating the height of a gas-phase transfer unit for vapors entering the primary condenser, H_{g1}, is not applicable for determining H_g. Because the final form of Equation 29-C, as it is shown in Section II-C-5 of Appendix C, is based on Figure 46 which is for flow of gases transverse to cylinders. The flow of residual vapors V_2 used in determining H_g were already past the condenser, therefore, the generalized Equation 29-C is not applicable any more. A sample calculation made in Appendix C, Section III-A-3, shows that H_g determined from Equation 29-C is about 6 times greater than those determined by either Equations 55 or 56.

The determination of the interfacial composition, which was subsequently used to obtain the gas-phase transfer data discussed above was based on the assumption that liquid-phase resistance to heat and mass transfer was absent. However, comparison of the interfacial composition and temperature with those for vapor and condensate had shown that, contrary to the assumption, both gas and liquid-film resistances were present. This can be confirmed further when the magnitudes of the individual gas-phase resistances are compared with the over-all gas-phase resistances for the three binary systems. Rather than individual data, the average resistances for each mixtures were determined and the ratios of $(1/k_g)$ / $(1/K_g)$ were determined. For the three binary systems these are summarized in Table XIV. This table shows that for most cases, as a matter of fact for ten out of twelve mixtures studied, the gas-phase resistance constituted less than 50 per cent of the total resistances, compared to 90 per cent reported (47) for distillation in wetted-wall

column. It is noticed from Table XIV that the liquid-phase resistance is predominant at both high and low concentrations of the more volatile component in the mixtures. It is further noticed that the higher the gas-phase resistance the higher the Murphree efficiency. Or in other words, as the liquid-phase resistance become significant less fractionation will take place, consequently the residual vapors become less enriched.

The number of over-all gas-phase transfer units for the three binary systems varied between 0.4 and 1.5, depending on the binary system and concentration. Figure 35 is a plot of Murphree plate efficiency, or degree of enrichment of residual vapors against the number of over-all gas-phase transfer unit. Again, rather than cluttering the chart with individual data, the average data for each mixture were plotted. The data for the three binary systems are represented with a single line. The Murphree plate efficiency which is used here to represent the degree of enrichment of residual vapors rises rapidly from 30 to 80 per cent as the number of over-all gas-phase transfer unit is increased from 0.4 to 1.5. Though experimental data are not available, from the trend of the curve it seems that many more over-all gas-phase transfer units are required to raise the efficiency from 80 to 100 per cent than were required to raise the efficiency from 30 to 80 per cent. From extrapolating the curve, one may find 6 to 10 transfer units are required to achieve 100 per cent efficiency.

The over-all height of a gas-phase transfer unit varied between 0.4 to 1.6 feet, depending on the binary system and composition.

Here, the over-all height of a gas-phase transfer unit would have

TABLE XIV

VARIATION OF RATIOS OF GAS-PHASE RESISTANCES WITH COMPOSITION AND MURPHREE PLATE EFFICIENCY

	Benzene-n-Heptane System				n-Heptane-Toluene System				n-Hexane-Toluene System			
Average y_{A2}, Mole Per Cent	85.1	75.9	63.9	39.7	86.7	67.8	53.0	35.8	85.8	78.7	65.9	31.6
Average $\frac{1/k_g}{1/K_g}$ (100)	28.5	40.0	42.3	18.2	29.6	68.4	26.6	31.7	37.5	48.0	52.6	23.3
Average Murphree Efficiency, Per Cent	65.6	75.4	75.6	43.3	40.5	73.8	35.8	49.0	45.7	46.6	56.8	33.0

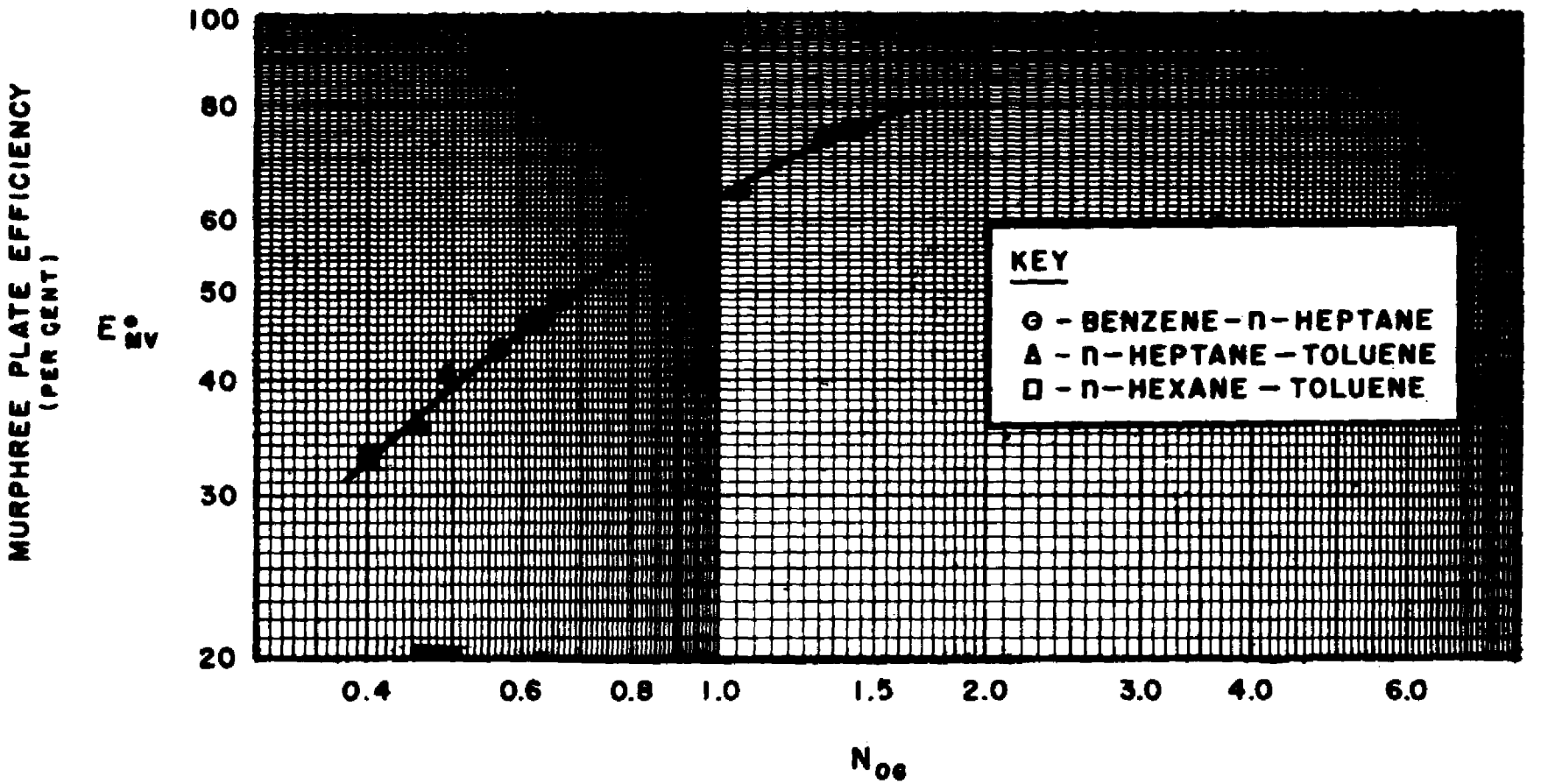


FIGURE 35. RELATIONSHIP BETWEEN NUMBER OF OVER-ALL GAS-PHASE TRANSFER UNIT AND MURPHREE PLATE EFFICIENCY

been the same as the height equivalent to a theoretical plate, H.E.T.P., if the equilibrium line were parallel to the operating line. The relationship between height of a transfer unit and height equivalent to a theoretical plate is given in the literature (73).

The H_{og} for the three binary systems were plotted against residual vapor composition. This is shown in Figure 36. Here again, rather than using the individual data, the average data for each mixture in a binary system were used. For both n-heptane-toluene and n-hexane-toluene systems a point fall above the curve. No explanation can be offered, but it is probably an error in data. For the three binary systems, H_{og} has a minimum value and becomes very large as the binary mixtures become concentrated in either component. The minimum point for the three binary mixtures occurs in the range of 65 to 70 mole per cent of the more volatile component in the mixtures. In a study on the distillation of binary and ternary mixtures in a wetted-wall column Qureshi and Smith (78) found out that the height of a gas-phase transfer unit depends on the concentration of the mixture in the column. Of the several mixtures studied by these authors, one is for n-heptane-toluene system, which their data are plotted in Figure 36. This data also show a minimum H_{og}. However, the curve flattens out in concentration range of 35 to 70 mole per cent n-heptane and the location of minimum point is in the neighborhood of 50 mole per cent. The authors give reference to other investigators that have obtained similar curves for distillation in packed and plate columns. Theoretically, based on two-film theory H_{og} should show a slight dependence on concentration, whereas experimental

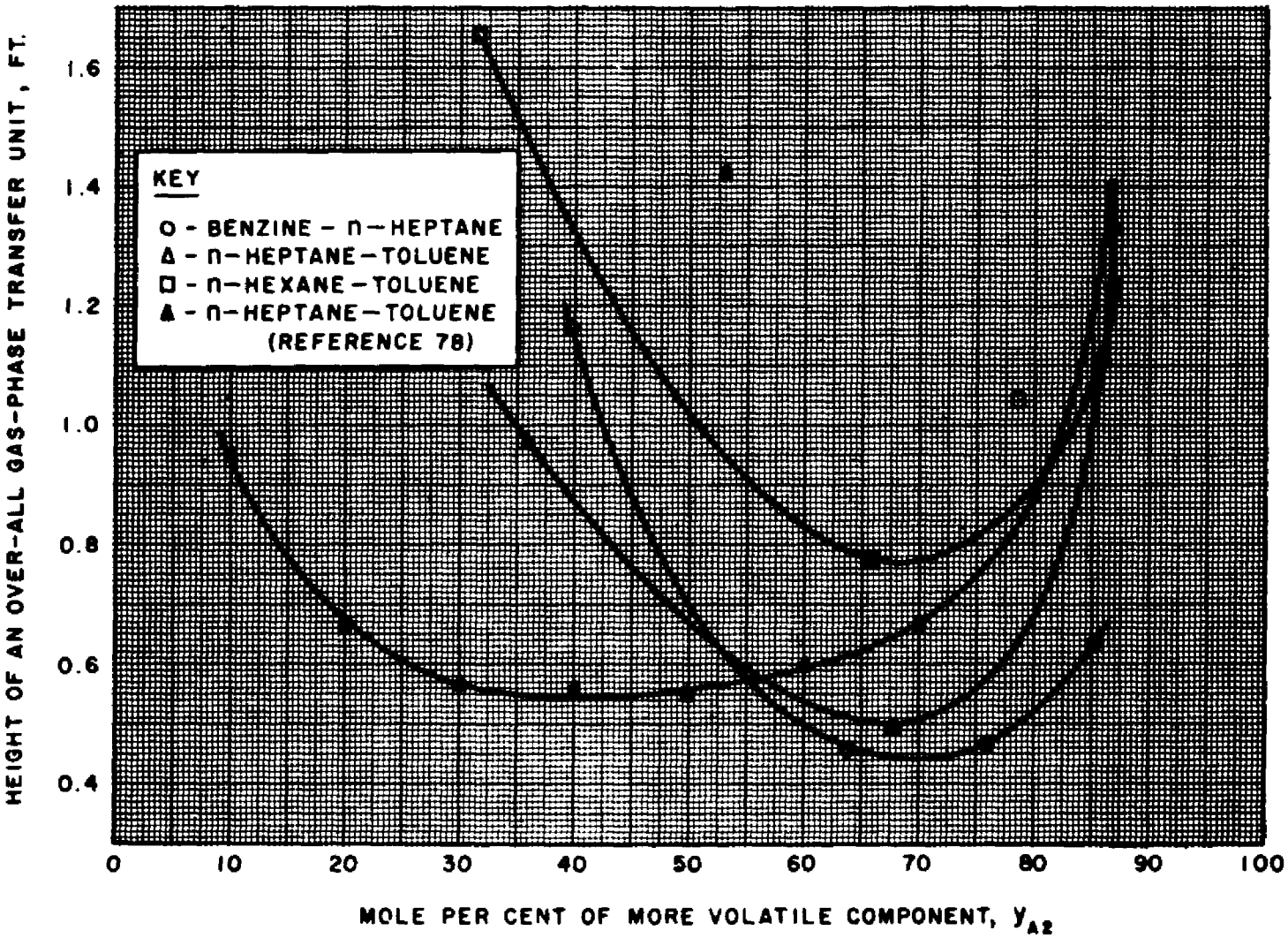


FIGURE 36. HEIGHT OF AN OVER-ALL GAS-PHASE TRANSFER UNIT FOR BINARY MIXTURES

data show otherwise. This is something that needs further investigation.

In a sample calculation liquid-phase mass transfer coefficient and the height of a liquid-phase transfer unit were calculated for Test No. 1, mixture A, of benzene-n-heptane system, Appendix C, Sections III-C-1 and III-C-2. Equations used for these determinations are applicable when the equilibrium line is straight over the region which includes the interfacial and bulk vapor compositions. The equilibrium line for benzene-n-heptane is straight over the region of compositions used in the sample calculation. The liquid-phase mass transfer coefficient was found to be 0.010 Lb.-moles/Hr.Ft.^{2} Mole Fraction, compared to 0.0265 Lb.-moles/Hr.Ft.^{2} Mole Fraction for the gas-phase mass transfer coefficient. The height of the liquid-phase transfer unit was found to be about 12 feet. It is felt that this is too high to be correct.

CHAPTER VII

CONCLUSIONS

The condensing heat transfer coefficients obtained in this work for pure hydrocarbons are about 20 to 30 per cent higher than the theoretical values predicted by Nusselt's equation. However, these coefficients are in good quantitative agreement with those reported in literature (62). Part of the difference between the experimental and theoretical coefficient is attributed to difference between the physical properties of the actual materials tested and the values reported for these in literature and used in calculating the theoretical values of the coefficient. The other part of the difference is believed to be due to turbulence induced in condensate film by vapor velocity or high drainage rate, even though the maximum vapor velocity in the jacket was less than one foot per second.

The result of this work on pure hydrocarbons indicated that the Nusselt's assumption that the condensate layer is in viscous flow is not entirely valid. It was found out that turbulence in condensate layer occurs at a much smaller value than the accepted critical value of 4200; most probably in the neighborhood of 250 to 300. At Reynolds number greater than 300, the heat transfer coefficient increases, rather than decreases, with increasing condensation rates. Based on mathematical and statistical analyses of the author's data for the pure hydrocarbons, the constants in Nusselt's equation can be changed to read

$$\frac{h_m}{\phi} = 3.6 \left(\frac{4\Gamma}{\mu_f} \right)^{-0.463}$$

Although this equation is limited to Reynolds number below 300, its standard error of estimate is 0.068.

The effect of non-linear temperature distribution and cross-flow within the film on condensing heat transfer coefficient as proposed in literature (10, 81) was studied for the four hydrocarbons. It was found out that the proposed correction does not increase the coefficient greatly. The maximum increase in the coefficient was about 2 per cent, which is negligible.

In condensation of binary vapors the material balance calculation on the unit showed that composition of the condensate from the primary condenser was almost the same as the original vapor entering the condenser, as required by the fact that the operation was such that condensation was almost total. The residual vapor (condensate from the auxiliary condenser) was consistently richer in the more volatile component than was the condensate from the primary condenser obtained at the same time. The McCabe-Thiele method of finding the number of theoretical plates, indicated that this unit is equivalent to a fraction of a theoretical plate. Using the Murphree plate efficiency equation, it was found out that for the three binary systems the degree of enrichment of the residual vapor varied between 30 to 80 per cent, depending on the binary system and composition.

With the assumption that the liquid-film resistance to mass transfer was absent, the interfacial temperature was found to be closer to the boiling point of the condensate than to either the dew point of the condensate or the vapor temperature. This is in accord with the work

of other investigators (75) that under condition of almost total condensation the boiling point of condensate can be taken as the interfacial temperature, if the liquid-phase resistance is not controlling. However, where there is significant change in composition across the liquid-film, the boiling point at the vapor-liquid interface would be different from that based on the average condensate composition; therefore, the boiling point of the condensate will not be the same as the interfacial temperature.

Film coefficients of heat transfer for benzene-n-heptane system can be obtained with about 10 per cent error by assuming a linear variation with mole fraction between the values of the pure components, when the temperature drop across the film is based on either the interfacial temperature or the boiling point of condensate. The deviation becomes much smaller at higher values of temperature drop across the film. The linearity of heat transfer coefficient with composition did not hold as true for n-heptane-toluene and n-hexane-toluene systems.

The sensible heat transferred either by bulk transport or by convection in cooling the vapor from its entrance temperature to the interfacial temperature is negligible and constitutes less than 3 per cent of the total heat transfer.

The experimental individual gas-phase mass transfer coefficients deviated an average of 10 per cent from those calculated using accepted correlations. Therefore, in the absence of experimental data, the gas-phase mass transfer coefficients may be determined from the appropriate correlation with a probable minimum average error of 10 per cent.

Mass transfer data indicated that resistance to heat and mass transfer existed in both vapor and liquid phases. For ten out of twelve binary mixtures studied, the liquid-phase resistance was more than 50 per cent of the total resistance. This is contrary to the assumption made in the previous studies made on this project (75) that the liquid-phase resistance to mass transfer was absent. Moreover, the liquid-phase resistance became predominant at high concentration of either component in the mixture. The fact that gas-phase resistance to mass transfer was small, as compared to the liquid-phase resistance, indicated that the major portion of mass transferred was by bodily transfer rather than by diffusion process. This conclusion is confirmed further when it is considered that the term C_o used in determining the sensible heat was large; consequently, the ratio of $C_o/1 - e^{-C_o}$ was essentially equal to C_o.

The enrichment of residual vapors is directly related to the extent that the gas-phase resistance is present. The higher the ratio of the individual gas-phase resistance to the over-all gas-phase resistance, the higher the degree of enrichment of residual vapors.

The number of over-all gas-phase transfer units varied between 0.4 and 1.5, depending on the binary system and the composition. The enrichment of residual vapors is also directly related to the number of over-all gas-phase transfer units. However, only at low values of enrichment did a given percentage increase in the number of over-all transfer units effect about the same percentage increase in the enrichment, whereas this was not true for high levels of enrichment.

A greater number of transfer units were required to raise the enrichment from 80 to 100 per cent than were required to raise it from 30 to 80 per cent.

The over-all height of a gas-phase transfer unit varied with the composition from 0.4 to 1.6 feet. For the three binary systems H_{og} appeared to have a minimum value and became very large as the binary mixtures became concentrated in either component. No theory could be presented to predict a minimum H_{og}.

CHAPTER VIII

SUGGESTIONS FOR FUTURE WORK

The present study on pure hydrocarbons has shown that the turbulence of the condensate film starts at a Reynolds number of about 250 to 300. Assuming that this is not due to the "drag" affect of the vapors which moved at a velocity of less than one-foot per second, the calculated Reynolds number is far less than 4200 which is suppoed to be a criterion for turbulence in the condensate film. As a part of heat transfer studies, further work should be done to establish the Reynolds number criterion for turbulence in the condensate film and its effect upon condensing heat transfer coefficient.

The present study may be extended further to determine heat and mass transfer coefficients for condensing binary hydrocarbons vapors composed of paraffinic and/or aromatic hydrocarbons. Such data are not available in literature.

As a part of a larger program in heat and/or mass transfer studies, in partial condensation the effects of fluid velocities and composition on the rate of heat and/or mass transfer and the enrichment of residual vapors may be investigated. Condensation with complete fractionation (equilibrium between vapor and liquid) and the fractionation effected in condensation may also be considered.

Mass transfer operation in condensing binary vapors should be studied further. The present investigation has shown that contrary to the accepted theory the liquid-phase resistance to mass transfer is appreciable and constitutes better than one-half of the total resistance.

This problem should be studied further and establish the limits under which it holds true. Should additional data verify this, liquid mass transfer data and relationships are required; the effect of the vapor and liquid diffusion on the interface composition should be studied. Johnstone and Pigford (47) found mass transfer rates in streamline flow region of a liquid film in a wetted column to be much higher than those predicted by theory. These authors report values for the heights of liquid-film transfer units, for distillation in a wetted-wall column one-half to one-quarter those predicted by theory for liquid-film Reynolds numbers below 500. This problem along with the effect of turbulence in the condensate film upon mass transfer rates may be considered in condensing binary vapors.

The relationship between H_{og} and composition and the minimum H_{og} shown here and elsewhere (78) require further investigation. No attempt was made in the present study to correlate the vapor and the liquid phase height of a transfer unit with the Reynolds number of the corresponding phases. However, such correlations may be developed and compared with the observed measurements shown by Kent and Pigford (50). For these studies, references (19,20,47,50,78,98) may be of value.

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A P P E N D I X

APPENDIX A

NOMENCLATURE

a - Interfacial area per unit volume of apparatus, Ft.^{2}/Ft.^{3}

A - Heat transfer area, Ft.^{2}

c - Concentration, Lb.-moles/Ft.^{3}

C_o - Defined by Equation 21.

C_p - Specific heat at constant pressure, Btu/Lb.°F., C_{pA} and C_{pB} are for components A and B. C_{pf}, liquid-film or gas-film specific heat. $C_p°$, Ideal state gas specific heat.

D - Diameter, Ft. D_1, Outside diameter of condensing bar, (3.908/12) Ft. D_2, Inside diameter of the jacket, (6.025/12) Ft. D_o, Outside diameter of pipe, Ft.

D_v - Diffusivity coefficient, Ft.^{2}/Hr.

e - Base of natural logrithm, 2.718...

$E°_{MV}$ - Murphree vapor efficiency, Equation 17-C, Appendix C

g - Acceleration due to gravity, 4.17×10^8 Ft/Hr.^{2}

g_c - Conversion Factor, 4.17×10^8 Lb. of Fluid. Ft./Hr.^{2} Lb. Force

G - Mass velocity of condensing vapor in the primary condenser, Lbs./Hr.Ft.^{2}. Based on surface area of condenser and rate of condensation in the primary condenser.

G' - Mass velocity of vapor, Lbs./Hr.Ft.^{2} used for calculating the Reynolds number of condensing vapor in the primary condenser. Based on one-half of cross sectional area of

flow and one-half of rate of condensation in the primary condenser.

G_1 - Mass velocity of vapor, Lbs./Hr.Ft.^{2} used for calculating the Reynolds number of vapor entering to the primary condenser. Based on one-half of cross sectional area of vapor flow and one-half of rate of flow of vapor entering the primary condenser.

G_M - Molar mass velocity of condensing vapor in the primary condenser, Lb-moles/Hr.Ft.^{2}. Based on one-half of cross sectional area of vapor flow and one-half of rate of condensation in the primary condenser.

G_{M1} - Molar mass velocity of vapor entering the primary condenser, Lb.-Moles/Hr.Ft.^{2}. Based on one-half of cross sectional area of vapor flow, one-half of rate of vapor flow entering the primary condenser, and molecular weight of entering vapor.

G_{M2} - Molar mass velocity of vapor leaving the primary condenser, Lb.-moles/Hr.Ft.^{2}. Based on one-half of cross sectional area of vapor flow, one-half of vapor flow rate leaving the primary condenser, and the molecular weight of vapor leaving the primary condenser.

h - Height that vapor is traveling in the jacket, Ft.

h - Heat transfer coefficient, Btu/Hr.Ft.^{2}°F. h_{BP}, based on boiling point of condensate. h_{DP}, based on dew point of condensate. h_I, based on interfacial temperature.

h_v, based on entering vapor temperature.

h_g - Gas-film coefficient of heat transfer, Btu/Hr.Ft.^{2}°F.

h_{gA}, h_{gB} - Enthalpy of saturated vapor of components A and B, respectively, Btu/Lb.

h_{lA}, h_{lB} - Enthalpy of saturated liquid of components A and B, respectively, Btu/Lb.

h_m - Mean coefficient of heat transfer, Btu/Hr.Ft.^{2}°F.

H_g - Height of an individual gas-phase transfer unit, based on the rate of condensation in the primary condenser, Ft.

H_{gl} - Height of an individual gas-phase transfer unit, based on the rate of vapor entering the primary condenser, Ft.

H_l - Height of an individual liquid-phase transfer unit, Ft.

H_{og} - Height of an over-all gas-phase transfer unit, Ft.

J_D, J_H - Dimensionless factors for mass and heat transfer, respectively.

k - Boltzmann Constant, gas constant/molecule,

1.381×10^{-16} cm.dyne/deg.

k - Thermal conductivity, Btu/Hr.Ft.^{2} °F/Ft. k_f thermal conductivity at the mean liquid-film temperature. k_A, k_B, thermal conductivity of components A and B in vapor phase. k_m thermal conductivity of vapor mixture.

k_g - Individual gas-phase mass transfer coefficient, Lb.-Mole/Hr.Ft.^{2} Mole Fraction.

k_l - Individual liquid-phase mass transfer coefficient, Lb.-moles/Hr.Ft.^{2} Mole Fraction.

K_g - Overall mass transfer coefficient, Lb.-Moles/Hr.Ft.^{2} Mole Fraction.

L - Effective condensing length or wall thickness, Ft.

L - Molar mass rate of liquid, Lb.-moles/Hr.

L_M - Molar mass velocity of liquid, Lb.-moles/Hr.Ft.^{2}

m - Slope of equilibrium curve, Equation 63.

M - Molecular weight, M_A for component A, M_B for component B.

n_D^{25} - Refractive index at 25°C. using sodium D line.

N_A, N_B - Rate of diffusion of components A and B, respectively, Lb.-moles/Hr.Ft.^{2}

N_g - Number of individual gas-phase transfer unit.

N_{og} - Number of overall gas-phase transfer units.

p - Partial pressure of diffusing vapor, Atmosphere. p_A and p_B are for components A and B. p_{AI}, partial pressure of component A at the interface. p_{AV}, partial pressure of component A in main body of vapor.

P - Total pressure, Atmosphere.

P_C - Critical pressure, Atmosphere.

Pr - Prandtl number, $C_p \mu/k$, a dimensionless term.

q - Heat transferred, Btu/Hr.

q_c - Heat flux per unit area arising from condensate cooling, Btu/Hr.Ft.^{2}

q_s - Heat flux per unit area received by condensate layer from vapor as sensible heat, Btu/Hr.Ft.^{2}

$q_{\lambda I}$ - Heat flux per unit area evolved by condensation at the

interface I-I, Btu/Hr.Ft.^{2}

Q - Total heat flux per unit area absorbed by the cooling medium, Btu/Hr.Ft.^{2}

r - Radius of pipe, Ft.

r - Distance of separation of molecules.

r_h - Hydraulic Radius, r_h = flow area/wetted perimeter, Ft.

R - Ratio of W/W_2.

Re.No. - Reynolds Number

S - Cross section of film of Condensate, Ft.^{2}

S - Sutherland Constant, $S = 1.47\ T_b$

Sc - Schmidt number $\mu/\rho D_v$, a dimensionless term.

t, T - Temperature, (not absolute, unless indicated, °F)

T_b - Normal boiling point, °K.

T_{BP} - Boiling point of condensate, °F.

t_c - Cooling water temperature, °F.

T_c - Critical temperature, °K.

T_{DP} - Dew point of the condensate, °F.

T_F - Mean film temperature, °F.

T_I - Temperature at the interface, °F.

T_r - Reduced temperature

T_V - Entering vapor temperature, °F.

T_W - Average wall temperature, °F.

v - Downward velocity of condensate film, Ft./Hr. v_m, mean condensate velocity, Ft./Hr.

V - Vapor flow rate, Lb.-moles/Hr. V_1, for vapor entering the

primary condenser. V_2, for vapor leaving the primary condenser (entering the auxiliary condenser).

V_b - Molecular volume at normal boiling point, cm.^{3}/gm.-mole. V_{bA} and V_{bB} are for components A and B, respectively.

V_c - Critical volume, Cm^3/gm.-mole.

w - Molar rate of condensation as used in Equation 31. Lb.-Mole/Hr.Ft.^{2}

W' - Mass rate of flow of condensate, Lbs./Hr.

W - Rate of condensation in the primary condenser, Gms./Min.

W_1, W_2 - Rate of vapor flow to the primary and the auxiliary condensers, respectively, Gms./Min.

x - Mole per cent or mole fraction in the liquid phase. x_A, for more volatile component in the primary condensate. x_B for less volatile. x_{A2}, for the auxiliary condensate. x_{AI}^* at the interface is in equilibrium with y_{AI}.

X - Weight per cent of a component in liquid phase. X_A, for more volatile component. X_B, for less volatile component.

y - Mole per cent or mole fraction in vapor phase. y_A, for more volatile component. y_B, for less volatile component. y_{A1} and y_{A2} refer to the more volatile component in vapors to the primary and the auxiliary condensers, y_A^* is in equilibrium with x_A. y_{AI} and y_{AV} refer to the more volatile component at the interface and in the main body of vapor, respectively.

Y - Weight per cent of a component in vapor phase. Y_{A1} and Y_{A2}

refer to the more volatile component in vapors entering and leaving the primary condensers.

z - Distance in direction of diffusion, Ft.

z - As defined by Colburn and Drew: Ratio of the rate of condensation of the more volatile component to the net rate of condensation, $N_A/(N_A + N_B)$.

Z - Thickness of the film or laminar layer, Ft. Z_l, for the liquid film. Z_g, for the vapor film.

Greek Letters:

α - Angle between any radius of a horizontal tube and a vertical plane at the axis, degree.

Γ - Total mass rate of flow over a surface per unit length.

Δ - Symbol for difference.

ϵ - Energy difference between the separated molecules and the molecules in the configuration in which they have the maximum energy of attraction, ergs.

f - Symbol for function of.

λ - Latent heat of condensation, Btu/Lb.

λ_I Latent heat of condensation at the interface, Btu/Lb.

λ' - Defined by Equation 12.

μ - Viscosity, Lbs/Hr.Ft.- μ_A, μ_B are viscosities of components A and B. μ_f Viscosity at the mean condensate-film temperature, T_f.

π - 3.1416 ...

ρ - Density, Lbs./Ft.^{3}. ρ_A and ρ_B are densities of components A and B. ρ_f, density of condensate film at T_f.

σ - Collision diameter for low velocity head-on-collisions, Angstrom.

ϕ - Nusselts physical property group, a dimensionless factor.

Θ_{12} - Defined by Equation B-8.

APPENDIX - B

PHYSICAL PROPERTIES AND OTHER DATA

In preference to experimentally determining the physical properties of the specific hydrocarbons used in this experiment, it was decided to used the best information available in literature. Although the reported properties may be highly accurate, they may not represent the properties of the particular grade of the hydrocarbons used for this investigation. With this limitation in mind, the following properties were either directly reported from the literature or were calculated from theoretical equations. When a property was reported by many investigators, the data were compared and the best data in this author's opinion (or the opinion of other authors who have compared the data of various investigators) were chosen to be used in calculations. When experimental data for a property of a hydrocarbon were not available in the literature or the available data were very meager and doubtful, a study was made of the theoretical or empirical equations proposed by various authors. The equations were compared against the experimental data for other hydrocarbons. Of these proposed equations, the one which deviated the least from the experimental data, was chosen to calculate the property in question.

I DENSITY

I A - Pure Liquids. The density of liquid hydrocarbons is given by Rossini (84) in the A.P.I. Project 44. The data in this project are reported to be very accurate. Therefore, no further search of literature

was made for this property.

I B - Pure Vapors. Vapor densities for benzene, n-heptane, and n-hexane are reported by Lange (55). No data could be found for toluene vapor.

An equation is presented by Dreisbach (28) for calculating the vapor densities. This method, which is based on a vast quantity of experimental data is believed to be very accurate, was used to calculate the toluene vapor density. As a check on Dreisbach's equation, the vapor densities of benzene, n-heptane, and n-hexane were also calculated and were compared with the data reported by Lange. For each of the three hydrocarbons, the calculated values were found to be very close to Lange's data. The maximum deviation from Lange's data for the three hydrocarbons was less than 0.5 per cent.

The equation proposed by Dreisbach is:

$$\log \rho = A^* - \frac{B^*}{t + C} \qquad \text{B-1}$$

up to temperature t_k, where

$$t_k = t_x + \frac{K}{c} \qquad \text{B-2}$$

$$t_x = (1.1\ T_c - 273.2) \qquad \text{B-3}$$

where T_c is the critical temperature, °K.

The constants A^*, B^*, C, c, K, t_k, and t_x for the four hydrocarbons are tabulated below:

	Benzene	n-Heptane	n-Hexane	Toluene
A^*	1.19411	1.31598	1.25366	1.27923
B^*	1127.9	1190.5	1097.44	1252.3
C	220.790	216.954	224.366	219.337
c	-0.13147	-0.14033	-0.14186	-0.11760
K	25	24	24	23
T_c, °K	562.6	540.0	507.9	594.0
t_x, °C	345.8	321.0	285.6	380.2
t_k, °C	155.6	150.0	117.0	184.6

The calculated data are summarized in Table XV.

As it was indicated earlier, Equation B-1, is accurate up to a temperature t_k, for each substance. Above the temperature t_k, other constants are introduced into the equation. However, since in all cases the temperatures at which it was desired to calculate the vapor densities were below t_k, Equation B-1 was satisfactory.

I C - Mixed Vapors. Knowing the densities of the pure components and the composition of each binary mixture, the vapor densities of each mixture were calculated. The densities of mixed vapors were used in calculating the Schmidt numbers of vapors entering the primary condenser. Therefore, in calculating the densities of mixed vapors, the composition of vapors entering the primary condenser were used rather than the composition of the condensates from the primary condenser. However, as shown in Table III, except for a very few cases, the differences between the composition of vapor entering the primary condenser and the condensate from the primary condenser is less than one mole per cent and very often less than 0.5 mole per cent. The calculated vapor densities are shown in Table XX and Figure 37 as a function of composition.

II VISCOSITY

II-A Pure Liquids. The viscosity of liquid hydrocarbons is given

TABLE XV

DENSITIES OF LIQUID AND VAPOR HYDROCARBONS

Liquids, Lbs./Ft.^{3}

Temperature °F.	Benzene	n-Heptane	n-Hexane	Toluene
60	55.89	43.50	42.01	55.02
80	55.16	42.92	41.41	54.37
100	54.44	42.33	40.78	53.74
120	53.70	41.75	40.15	53.09
140	52.94	41.16	39.49	52.45
160	52.18	40.53	38.79	51.81
180	51.44	39.87	38.20	51.17
200	50.62	39.22		50.53
220		38.55		49.89
				49.25

Vapors, (Lbs./Ft.^{3}) x 10^3

Benzene				n-Heptane				n-Hexane				Toluene	
Calculated		Lange		Calculated		Lange		Calculated		Lange		Calculated	
t, °F.	ρ	t, °F.	ρ	t, °F.	ρ	t, °F.	ρ	t, °F.	ρ	t, °F.	ρ	t, °F.	ρ
68	20.20	194	224.75	68	12.23			68	36.15	194	362.09	68	7.13
86	31.05	212	293.42	86	19.53	212	224.75	86	54.31	212	468.22	86	11.30
104	46.19	230	374.58	104	30.08	230	293.42	104	79.03	230	599.33	104	17.87
122	66.74	248	480.71	122	44.85	248	380.82	122	112.00	248	749.16	122	26.66
140	93.90	266	599.33	140	64.99	266	486.95	140	154.82	266	936.45	140	39.10
158	129.04	284	736.67	158	91.77	284	611.81	158	209.39	284	1167.44	158	55.86
176	173.68	302	898.99	176	126.58	302	761.64	176	278.25	302	1435.89	176	77.91
194	229.28			194	171.00			194	361.47			194	106.38
212	297.54			212	226.61			212	463.17			212	142.40
230	380.00			230	295.23			230	584.65			230	187.35
248	476.46			248	378.67			248	727.93			248	242.48

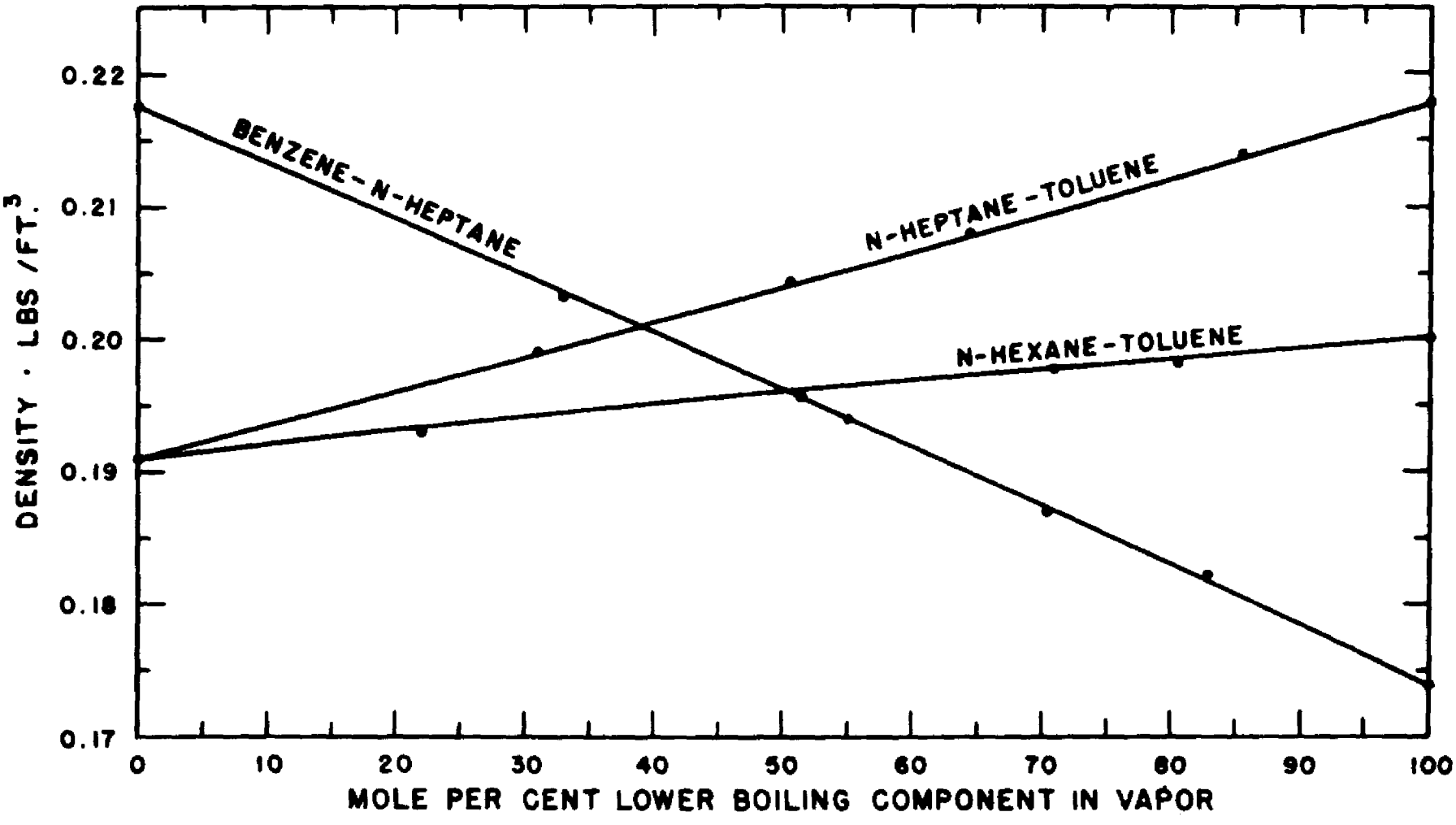

FIGURE 37. DENSITY OF BINARY VAPOR MIXTURES

in A.P.I. Project 44 (84). The viscosity data in this project are reportedly very accurate. Therefore, no further search of literature was made for this property. The data in the A.P.I. Project 44 are given in metric units. For this work they were converted to English units.

II-B Pure Vapors. Viscosity for benzene vapor are reported by Nasini (70), Titani (102) Uyehara and Watson (108), Craven and Lambert (24), and McCoubrey, et al. (63). The data from different sources are remarkably close together. Vapor viscosity for n-heptane is reported by Melaven and Mack (64) and Lambert and his co-workers (54). The data of Melaven and Mack are very old and are about 8 per cent lower than those reported by Lambert, et al. Vapor viscosity values for n-hexane are reported by Titani (102), McCoubrey, et al. (63), Lambert, et al. (54) and recently by Thorne (101). For the given temperature ranges, the data of Titani, Lambert, et al., and Thorne checks very closely. McCaubrey's data are about 3.5 per cent lower than those by Titani, Lambert, et al., and Thorne. Only one set of experimental data was found for toluene vapor, that reported by Nasini.

All of the above data were given in metric units. The data were converted to the engineering units and are shown in Table XVII.

Since the experimental data for n-heptane and toluene were limited, a survey of the literature was made to find a viscosity-temperature correlation which would give the most accurate data. The purpose of finding such a correlation was to check it against the available experimental data for benzene and n-hexane and find out how accurate it is. Should an accurate correlation be located, then the viscosity of

n-heptane and toluene vapors would be calculated and compared with the reported experimental data to see how closely the experimental data checks with the correlation.

Among the many viscosity - temperature relationships proposed by various investigators (1,8,41-44,57,100) for calculating the gas viscosities, the following equation is reported (8,79) to give viscosity data closer to experimental values than any other correlation.

$$\mu = 3.33 \times 10^{-3} \frac{(M\ T_c)^{0.5}}{V_c^{2/3}} \quad f \quad \frac{k\ T}{\epsilon} \qquad \text{B-4}$$

This is a modification of an equation originally proposed by Hirschfelder and his co-workers (41-43). The term ϵ/k which is a characteristic constant for each gas for 45 common gases are reported in literature (87, 41-44, 79). For other gases Hirschfelder, et al., recommends the approximate estimation of ϵ/k by the following empirical relationships:

$$\epsilon/k = 0.75\ T_c = 1.33\ T_r \qquad \text{B-5}$$

$$\epsilon/k = 1.39\ T_b \qquad \text{B-6}$$

In a later work the authors (44) revised the constants 0.75 and 1.39 in the above equations to 0.77 and 1.15, respectively. Reportedly (8,44) Equation B-5 is preferable to Equation B-6.

Recently Reid and Sherwood (79) compared the experimental viscosity data for 31 non-organic, organic, and hydrocarbon vapors, with calculated values using Equation B-4, and other viscosity-temperature relationships available in literaaure. They found out that for the 31 gases compared,

the data calculated by Equation B-4 deviated about 3 per cent from experimental data, as compared to 5 to 6 per cent deviation obtained when using other correlations (1,8,57). Among the many gases they compared the experimental data with the calculated data are benzene, n-hexane, and toluene which are of interest in this work. The experimental viscosity data for these hydrocarbons are compared with the calculated values in Table XVI. The comparison shows that Bromley and Wilke (8) correlation, Equation B-4, gives less per cent error than the other correlations.

Based on this review, Equation B-4 was used to calculate the vapor viscosities of benzene, n-heptane, n-hexane, and toluene. It was mentioned before that in a later work Hirschfelder et al., (44) revised the constant 0.75 in Equation B-5 to 0.77. Using this new constant would give a factor of 0.003376 for Equation B-4, rather than 0.003330, as proposed by Bromley and Wilke. The new factor 0.003376 was used in calculating the vapor viscosities. The calculated viscosities are shown in Table XVII. The data for benzene was found to be remarkably close to the experimental data by various investigators and indicated the accuracy of the equation. The calculated data for n-hexane were very close to Lambert, et al., (54) and fell on the same line with Thorne (101) and Titani (102) data. The calculated data for n-heptane differed by less than 0.3 of a per cent from those reported by Lambert, et al. The experimental data for toluene vapor by Nasini (70) were 2.3 per cent greater than the calculated data. The deviation being so small, it was preferred to use the experimental rather than calculated data, as it was done for benzene,

TABLE XVI

COMPARISON BETWEEN CALCULATED AND EXPERIMENTAL VALUES OF GAS VISCOSITY

Values of Viscosity in Centipoises x 100

Hydrocarbon	Temperature, °C.	Experimental Viscosity	Values Calculated By Methods Of: Licht and Stechert (57)		Arnold (1)		Bromley and Wilke (10): Equation B - 4		Bromley and Wilke (10): Equation 19 in Reference (8)	
			Calculated	% Error [2]	Calculated	% Error	Calculated	% Error	Calculated	% Error
Benzene	20.0	0.746	0.738	-1.0	0.697	-6.6	0.746	0	0.755	+1.2
	77.8	0.870	0.896	+3.0	0.853	-2.0	0.901	+3.5	0.911	+4.7
	150.0	1.070	1.088	+1.7	1.044	-2.4	1.082	+1.1	1.098	+2.6
	250.0	1.311	1.342	+2.4	1.300	-0.8	1.338	+2.1	1.350	+3.0
Average				2.0		3.0		1.7		2.9
n-Hexane	35.0	0.709	0.665	-6.2	0.619	-13.0	0.693	-2.3	0.673	-5.1
	77.8	0.795	0.762	-4.2	0.715	-10.0	0.793	-0.3	0.770	-3.2
	150.0	0.934	0.925	-1.0	0.870	-6.9	0.958	+2.6	0.930	-0.4
	225.0	1.098	1.079	-1.7	1.027	-6.5	1.122	+2.2	1.089	-0.8
Average				3.3		9.1		1.9		2.4
Toluene	60.0	0.789	0.770	-2.4	0.726	-7.9	0.791	+0.3	0.786	-0.4
	150.0	1.008	0.993	-1.4	0.948	-5.9	1.011	+0.3	1.005	-0.3
	250.0	1.227	1.230	+0.2	1.181	-3.7	1.247	+1.6	1.240	+1.0
				1.3		5.8		0.7		0.6

[1] Courtesy of "The Properties of Gases and Liquids" by Reid and Sherwood (79), copyright 1958, McGraw-Hill Book Company.

[2] $\% \text{ Error} = \frac{\text{calculated - experimental}}{\text{experimental}} \times 100$

TABLE XVII

VISCOSITIES OF LIQUID AND VAPOR HYDROCARBONS

Units: Liquids, (Lbs./Hr. Ft.) x 10^5
Vapors, (Lbs./Hr. Ft.) x 10^4

I BENZENE

Liquid		Vapor											
		Calculated		Craven & Lambert		McCoubrey		Nasini		Titani		Uyehara & Watson	
t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ
40	1,975	170	215.2	95	192.6	72	180.2	60	178.1	56	178.6	212	229.2
60	1,680	180	218.4	122	200.6	129	204.8	106	194.1	122	222.1	322	267.9
80	1,425	190	222.1	149	207.9	214	230.7	213	227.7	268	249.5	431	306.3
100	1,226	200	225.8	172	210.5	271	249.2	301	259.9	322	268.6	536	339.7
120	1,070	210	229.1	185	222.6	352	274.5	387	284.6	382	289.9	595	359.1
140	944	220	232.6					479	316.8	486	325.0		
160	840	230	236.0							595	359.1		
180	752												

II n-HEPTANE

Liquid		Vapor					
		Calculated		Lambert, et al.		Malaven & Mack	
t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ
40	1,207	170	176.8	149	172.3	213	173.5
60	1,061	180	179.8	162	175.2	303	197.9
80	944	190	182.6	176	179.5	396	223.1
100	847	200	185.5	195	183.7	485	261.3
120	762	210	188.2				
140	694	220	191.2				
160	632	230	194.1				
180	576	240	197.0				
200	531						

III TOLUENE

Liquid		Vapor			
		Calculated		Nasini	
t, °F.	μ	t, °F.	μ	t, °F.	μ
40	1,754	200	191.1	141	190.9
60	1,503	210	194.2	205	213.2
80	1,309	220	197.0	305	244.9
100	1,153	230	200.2	382	268.1
120	1,025	240	203.1	483	298.1
140	919	250	206.0		
160	830	260	209.2		
180	753	270	212.0		
200	687	280	214.9		
220	630	290	217.9		
		300	220.9		

IV n-HEXANE

Liquid		Vapor									
		Calculated		Lambert, et al.		McCoubrey		Titani		Thorne	
t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ	t, °F.	μ
40	880	120	170.8	122	172.8	60	149.0	250	209.6	122	171.8
60	791	140	176.9	140	177.6	169	178.5	322	231.8	140	176.6
80	708	160	182.8	149	179.6	217	192.7	372	247.1	176	187.6
100	645	180	188.8	172	187.6	252	202.0			212	198.4
120	583	200	194.8			302	218.9			248	209.3
140	538	220	200.8			372	237.4			284	220.2
160	484	240	206.9							320	229.9
		260	212.8								
		280	218.8								
		300	224.7								

n-heptane, and n-hexane.

II C - Mixed Vapors. Based on modern molecular theory, Hirschfelder and several co-authors (41-44) have developed relations for calculation of viscosity of non-polar gas mixtures at low pressure. These relations are based on quantitative evaluation of intermolecular forces and the use of collision integrals. The relations reproduce the experimental data with high degree of precision; however, they are quite complicated and cumbersome to use. Less complex, though less accurate, relation is given by Buddenberg and Wilke (15) which later was modified by Wilke (114). This equation is presented

$$\mu_m = \frac{\mu_1}{1 + \frac{y_2}{y_1}\Theta_{12}} + \frac{\mu_2}{1 + \frac{y_1}{y_2}\Theta_{21}} \qquad \text{B-7}$$

$$\Theta_{12} = \frac{\left\{1 + \left[(\mu_1/\mu_2)(\rho_2/\rho_1)\right]^{0.5} (M_1/M_2)^{0.25}\right\}^2}{(4\sqrt{2})\ (1 + M_1/M_2)^{0.5}} \qquad \text{B-8}$$

Here μ, ρ, y, and M are viscosity, density, mole fraction, and molecular weight. Subscripts 1 and 2 refer to components 1 and 2. To obtain Θ_{21} the subscripts are interchanged.

Wilke compared 17 sets of experimental data for binary systems with those calculated with Equation B-7 and reports the equation deviated an average of less than 1 per cent from the experimental data. He also extended these relations for multicomponent mixtures and found these relations to represent data on viscosities of a variety of multicomponent

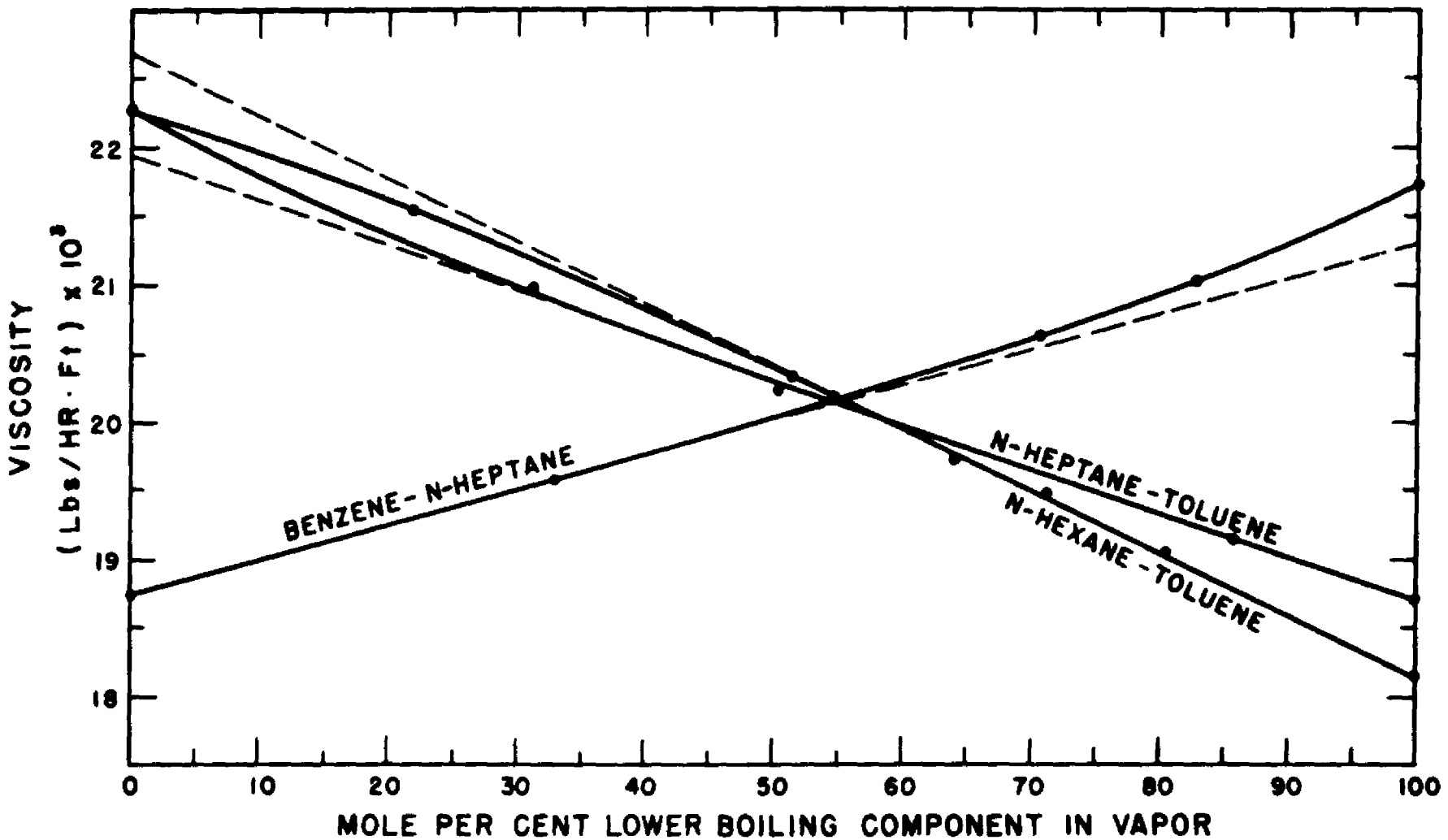

FIGURE 38. VISCOSITY OF BINARY VAPOR MIXTURES

gas mixtures with an average deviation of less than two per cent.

The viscosities of the binary mixtures for benzene-n-heptane, n-heptane-toluene, and n-hexane-toluene systems were calculated by using Equation B-7. For each binary mixture, the vapor compositions were taken as the average composition of the incoming vapors to the primary condenser.

The calculated viscosity data are shown in Table XX and are plotted as a function of composition of incoming vapor to the primary condenser. As Figure 38 shows, the lines of viscosity against composition have slight curvature and surprisingly in each case the curvature of the line starts when vapor composition is predominantly aromatic. It is very doubtful if this is of any significance. The maximum per cent deviations from straight line are +1.97, +1.43, and -2.06 per cents for benzene-n-heptane, n-heptane-toluene, and n-hexane-toluene systems, respectively.

III VAPOR DISSUFION COEFFICIENTS

Wilke and Lee (115) have reviewed critically the existing methods for estimating the diffusion coefficients of gases and vapors. The authors have compared the experimental data with correlations presented by Arnold (1), Gilliland (37), and Hirschfelder and his co-authors (41-44). Bird (5) has presented a comprehensive study and review on theory of diffusion. Recently Reid and Sherwood (79) reviewed the correlations presented by the above authors as well as an empirical equation by Slattery and Bird (96) and tested each of the equations by comparison with experimental data.

Comparing experimental data with those obtained from Arnold,

Gilliland, and Hirschfelder, et al., correlations Wilke and Lee concluded that Hirschfelder, et al., correlation would give the most accurate diffusion data for pairs of non-polar gases. This equation is expressed as

$$D_v = 9.2916 \times 10^{-4} \frac{\left(\frac{1}{M_1} + \frac{1}{M_2} \right)^{0.5}}{P (\sigma_{12})^2 \left[f (k T/\epsilon_{12}) \right]} \qquad \text{B-9}$$

Hirschfelder, et al., (41) point out that the force constants σ and ϵ/k are best obtained from viscosity data; the values obtained in this manner for a variety of gases are listed in references (8,41-44, 79, 115). It was pointed out in Section II-B that ϵ/k for other gases may be estimated from either Equation 5-B or 6-B.

The term σ may be estimated from the following empirical relationship (47).

$$\sigma = 0.833 \, V_c^{1/3} \qquad \text{B-10}$$

For a binary system, σ_{12} is taken as the arithetic mean of σ_1, and σ_2 and ϵ_{12} as the geometric mean of ϵ_1 and ϵ_2.

$$\sigma_{12} = \frac{\sigma_1 + \sigma_2}{2} \qquad \text{B-11}$$

$$\epsilon_{12} = (\epsilon_1 + \epsilon_2)^{0.5} \qquad \text{B-12}$$

Wilke and Lee (115) compared 64 sets of selected experimental data with the calculated values from Arnold, Gilliland, and Hirschfelder equations. The average and maximum per cent deviations from experimental

data for these equations are tabulated below:

Method	Per Cent Deviation	
	Average	Maximum
Arnold	8.4	20.5
Gilliland	20.0	46.8
Hirschfelder (σ and ϵ/k from viscosity data)	7.0	21.4

The authors also found that the average and maximum per cent error in Hirschfelder equation can be reduced from 7 and 21.4 to 3.9 and 16.0 per cents, respectively, if the constant 9.2916 in Equation B-9 be replaced with

$$10.7 - 2.46 \left(\frac{M_1 + M_2}{M_1 M_2}\right)^{0.5} \qquad \text{B-13}$$

where M_1 and M_2 are molecular weights of the components. There is an error in the original reference (115), where it is proposed the constant 9.2916×10^{-4} be replaced by Equation B-13.

Reid and Sherwood (79) compared and evaluated the several methods of estimating diffusion coefficients with 80 sets of experimental data, accumulated from a variety of sources, for different binary gas systems. They compared the experimental data with Arnold, Gilliland, Hirschfelder--where the force constants were determined from viscosity data and from empirical equations--and Slattery and Bird equations. The authors tabulated D_vP from experimental data against those calculated with above equations. The term D_vP is the product of D_v in square centimeters per second and pressure in atmospheres. Their average and maximum per cent

deviations from experimental for correlations proposed by various authors is given below:

Method of Calculating Diffusion Coefficients	Per Cent Deviation From Experimental Data	
	Average	Minimum
Arnold	11	41
Gilliland	16	82
Hirschfelder		
Force constants from viscosity data	6	19
Force constants estimated	10	33
Slattery and Bird	10	45

It is noticeable from the preceding tables that the most erratic is Gilliland equation. Arnold, Hirschfelder--when the force constants are estimated--and Slattery and Bird equations have about the same average per cent error. It is evident that Hirschfelder equation, when the force constants are determined from viscosity data, is the most accurate of all others.

Upon this basis, Hirschfelder equation was used to calculate the diffusion coefficients for the three binary systems tested in this investigation. The correction factor, Equation B-13, proposed by Wilke and Lee was introduced in these calculations. For the three binary systems under study, the magnitude of Equation B-13 as compared to the constant 9.2916 given by Hirschfelder is small, as shown below:

Binary System	Magnitude of Equation B-13	Ratio of Equation B-13 to 9.2916
Benzene-n-heptane	10.3287	1.1116
n-heptane-toluene	10.3450	1.1133
n-hexane-toluene	10.3314	1.1119

As the tabulated ratios show, the Wilke and Lee correction factor would increase the diffusion coefficients calculated by Equation B-9 about 11 per cent. The calculated diffusion coefficients for the three binary systems are shown in Table XX.

IV SCHMIDT NUMBERS

Knowing the required properties for each binary vapor mixture, the Schmidt numbers and $(Sc)^{0.56}$ were calculated. These are tabulated in Table XX.

V SPECIFIC HEAT

V-A Pure Liquids. Experimentally determined specific heat at constant pressure for benzene and toluene are reported by Burlew (16). Data for n-hexane are given by Douslin (27), and Connolly, et al. (23), and for n-heptane by Helfrey, et al. (39) and Douglas, et al. (26). Sakiadis and Coates (88,89) have presented two methods to calculate specific heat of liquids. In their first method the authors presented a theoretical equation for predicting the specific heat at constant volume and constant pressure of organic liquids as a function of temperature from the data on the velocity of sound, infrared and Raman spectra. Using this correlation the authors found that the average and

maximum deviations for 100 organic liquids at 68°F were 1.5 and +4.0 per cents, respectively. Deviations for seven organic liquids at a total of 50 temperatures, ranging from -184°F to 266°F, were 2.44 per cent average and +7.7 per cent maximum. Agreement was best at high temperatures.

The second method of Sakiadis and Coates, which was developed for hydrocarbon liquids, is based on a modified statement of the theory of corresponding states. The correlation developed by this method was tested by the authors against experimental data for 100 points representing a wide temperature range. The average and maximum deviations were found to be 0.9 and -2.7 per cents, respectively.

For this work the specific heat of liquid benzene, n-heptane, n-hexane, and toluene were calculated by the second method of Sakiadis and Coates. The calculated data were compared with the experimental data reported earlier. In all cases the data determined by Sakiadis and Coates' method were found to be remarkable close to the experimental data, with the maximum deviation being less than two per cent. All of these data are summarized in Table XVIII and shown graphically in Figures 39 through 42.

Several other methods for predicting the specific heat of liquids are available in literature. These methods, as reviewed by Gambill (34) and Reid and Sherwood (79), are less accurate than Sakiadis and Coates' methods; therefore, these methods were not reviewed.

V-B Pure Vapors. The vapor specific heat at one atmosphere for benzene are reported by Scott, et al. (93), Montgomery (66), and Pitzer (74).

Waddington, et al. (111) determined specific heat at one atmosphere for n-heptane, and Waddington and Douslin (112) determined the atmospheric specific heat for n-hexane. Meager data of Montgomery (only at two temperatures) (66) and Pitzer (74) data (three points at different pressures) are the only experimental data available for toluene vapors. Specific heat data at zero pressure, ideal gas state, for the four hydrocarbons used in this investigation are also given by Rossini (84) in A. P. I. Project 44. These data are also summarized in Table XVIII and shown graphically in Figures 39 through 42.

The data for benzene by Scott, et al., Montgomery, and Pitzer are very close to each other and as shown in Figure 39 they all fall on the same line or are very close to it. The data of Montgomery and Pitzer for toluene vapor are scattered. The main reason is that two out of three points by Pitzer were determined at pressures lower than one atmosphere. Being uncertain about which data to use, it was decided to use Edmister (29) method, calculate the specific heat for vapor toluene at one atmosphere and compare the calculated data with those of Montgomery and Pitzer. The two points by Montgomery are about 2 per cent higher than Edmister data. Pitzer data at one atmosphere fall on the Edmister plot of specific heat versus temperature and the other two points, at 0.53 and 0.22 atmospheres deviate only 0.2 and 0.8 per cents, respectively, from Edmister data. On this basis Montgomery data for toluene vapor were not used in this project.

Rossini data at zero pressure, for the four hydrocarbons, are slightly lower than the data reported for atmospheric pressure. The

TABLE XVIII

SPECIFIC HEAT OF LIQUID AND VAPOR HYDROCARBONS

Units: (Btu/Lb. °F.) x 10^3

I [illegible]

Liquid				Vapor							
Burlew		Sakiadis and Coates		Montgomery		Pitzer		Smith et al.		Rossini	
t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p° [1]
50	404.0	40	404.0	206	325.6	239	336.7	160	300.0	68	264.8
68	410.7	60	408.0	242	343.2	274	341.8	208	327.7	100	281.9
86	418.2	80	415.0	278	356.6	291	363.6	244	353.2	140	293.0
104	426.3	100	424.0			311	366.1	325	380.6	200	313.0
122	434.5	120	432.0			374	391.7	388	407.3	260	333.0
140	442.7	140	441.0			406	404.5			270	346.0
158	450.4	160	451.0							300	360.4
176	457.3	180	462.5							320	370.0

II n-HEPTANE

Liquid						Vapor			
Douglas et al.		Halfrey et al.		Sakiadis and Coates		Waddington et al.		Rossini	
t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p°
46	520.5	80	531.8	80	535.0	183	471.2	68	390.2
62	528.9	100	541.4	100	543.0	212	485.6	100	409.3
77	536.1	120	551.2	120	554.0	261	509.8	200	468.4
98	546.9	140	561.3	140	564.0	322	541.9	220	479.5
116	556.4	160	571.8	160	577.0	379	571.5	240	491.9
134	566.2	180	582.8	180	590.0			260	503.0
152	576.4	200	594.5	200	604.0			280	514.0
170	587.0	220	606.9	220	618.0			300	525.0
188	597.8							320	536.0
206	608.9							340	547.0
224	620.4							360	558.0

III n-HEXANE

Liquid						Vapor			
Douglin		Connolly et al.		Sakiadis and Coates		Waddington and Douglin		Rossini	
t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p°
41	524.3	80	543.5	80	539.0	141	468.8	100	410.4
48	528.2	100	555.9	100	550.0	197	476.2	180	450.0
64	535.5	120	569.0	120	562.0	258	508.6	220	473.0
82	544.7	140	582.5	140	575.0	321	542.2	260	495.0
		160	596.3	160	588.0	384	576.7	280	506.0
		180	610.6	180	601.0			300	518.0
		200	626.1	200	617.0			340	540.0

IV TOLUENE

Liquid				Vapor					
Burlew		Sakiadis and Coates		Montgomery		Calculated [2]		Rossini	
t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p	t, °F.	C_p°
41	396.7	40	395.5	242	368.2	190	334.9	60	359.9
50	398.7	60	404.0	278	383.7	210	344.3	77	369.0
68	404.1	80	412.0	Pitzer		260	358.4	100	380.9
86	410.9	100	423.0	t, °F.	C_p	260	367.3	180	395.5
104	418.8	120	420.0			290	380.8	200	335.0
122	427.5	140	437.0	248	358.2	340	402.6	240	350.0
140	436.7	160	445.5	311	390.7			260	359.2
158	446.0	180	456.0	374	415.6			280	369.0
176	455.2	200	466.0					300	378.5
194	463.9	220	477.0					320	388.5
212	471.8							340	397.0
230	478.6							360	406.0

[1] For the ideal gas state.

[2] Calculated by the Edminster's method.

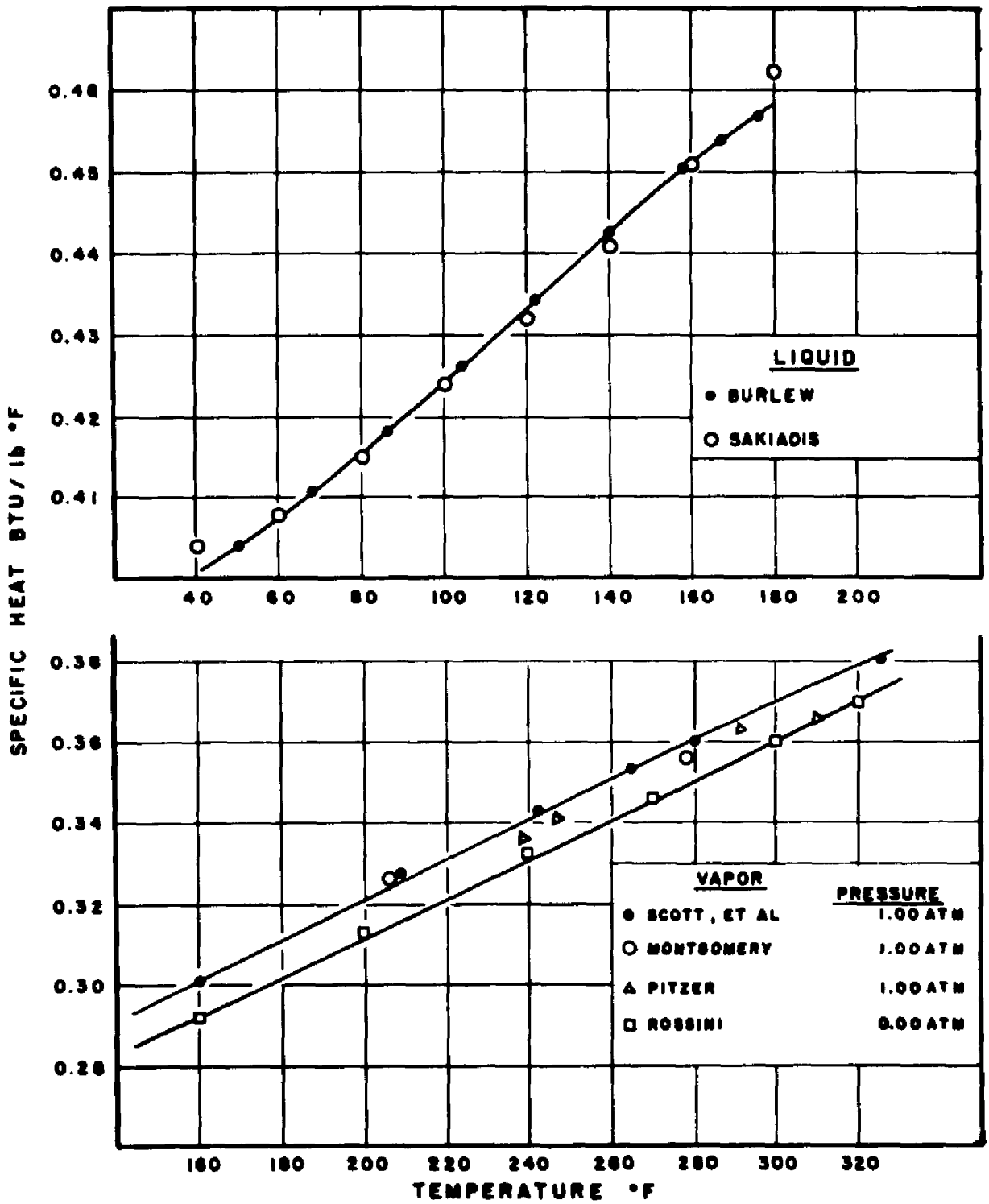

FIGURE 39. SPECIFIC HEAT OF LIQUID AND VAPOR BENZENE AT ONE ATMOSPHERE

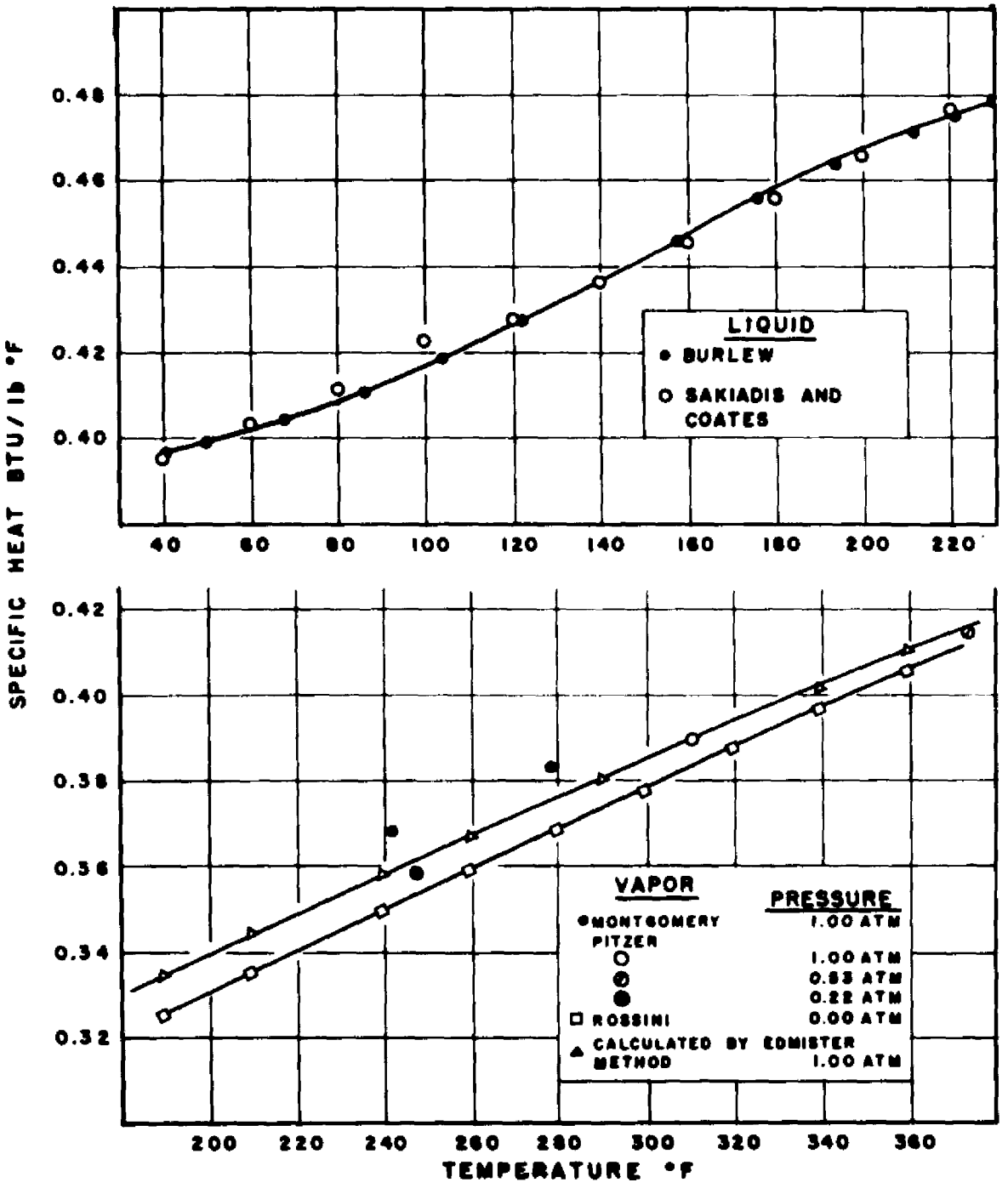

FIGURE 40. SPECIFIC HEAT OF LIQUID AND VAPOR TOLUENE AT ONE ATMOSPHERE

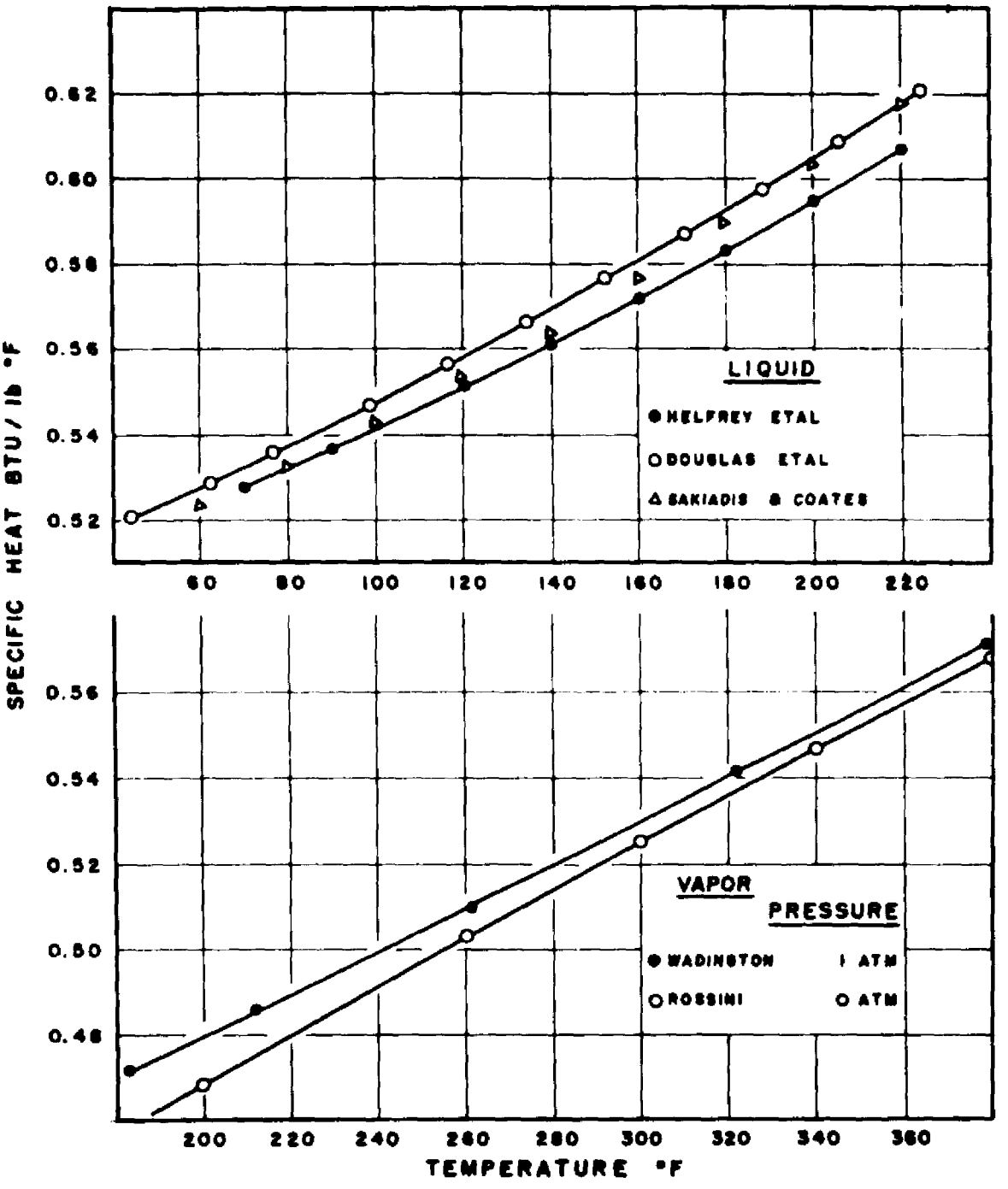

FIGURE 41. SPECIFIC HEAT OF LIQUID AND VAPOR n-HEPTANE AT ONE ATMOSPHERE

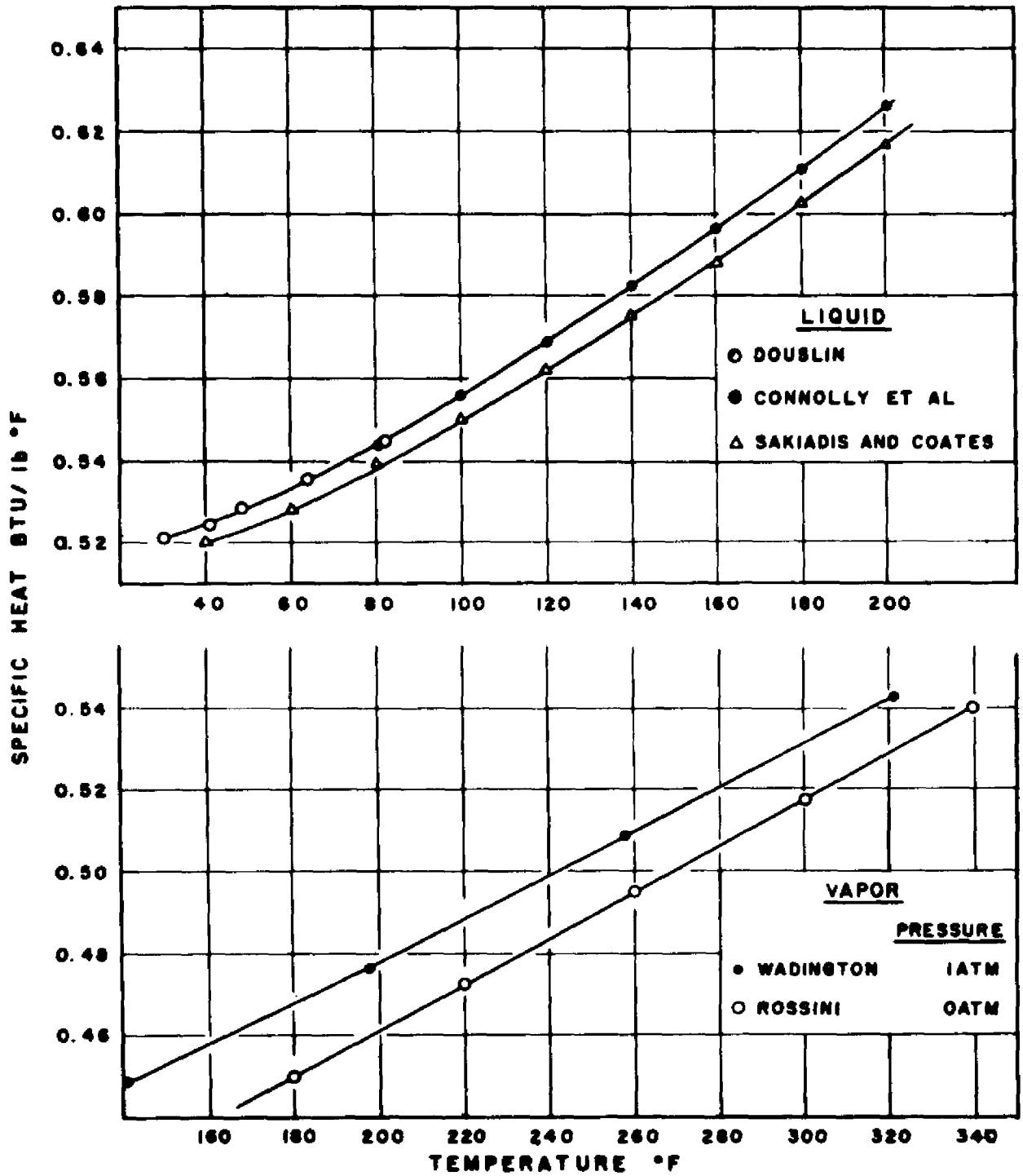

FIGURE 42. SPECIFIC HEAT OF LIQUID AND VAPOR n-HEXANE AT ONE ATMOSPHERE

maximum difference between zero and atmospheric pressure specific heats of the four hydrocarbons is less than 3 per cent - for n-hexane. This indicates that in the absence of experimental data at one atmosphere, the vapor specific heats at zero pressure, as are given in A.P.I. Project 44, can be used without any serious error.

V-C Specific Heat of Vapor Mixtures. The specific heat for the four mixtures in each of the three binary systems studied were calculated. These data are summarized in Table XX and are shown graphically as a function of composition in Figure 43. For each system the plot of specific heat versus composition deviated slightly from a straight line relationship. In the same manner as viscosity of vapor mixtures, the deviation from straight line occurred when the composition of the vapor was predominantly aromatic. The maximum deviation from straight line was -5.6 per cent for benzene-n-heptane, +1.8 per cent for n-heptane-toluene, and +1.0 per cent for n-hexane-toluene system.

VI THERMAL CONDUCTIVITY

VI-A Pure Liquids. Based on a comprehensive literature survey (85,86) on thermal conductivity of liquids and their own experimental data, Sakiadis and Coates (87,90) have presented two fairly complex methods for predicting the thermal conductivity of liquids. These methods are presented and discussed in great detail by the authors and reviewed by Gambill (32) and Reid and Sherwood (79). Of the two methods, one is based on a model of the liquid state in which the molecules are assumed to be oriented in isothermal chain which transfer energy (heat) through interactions with other chains not in their isothermal surface.

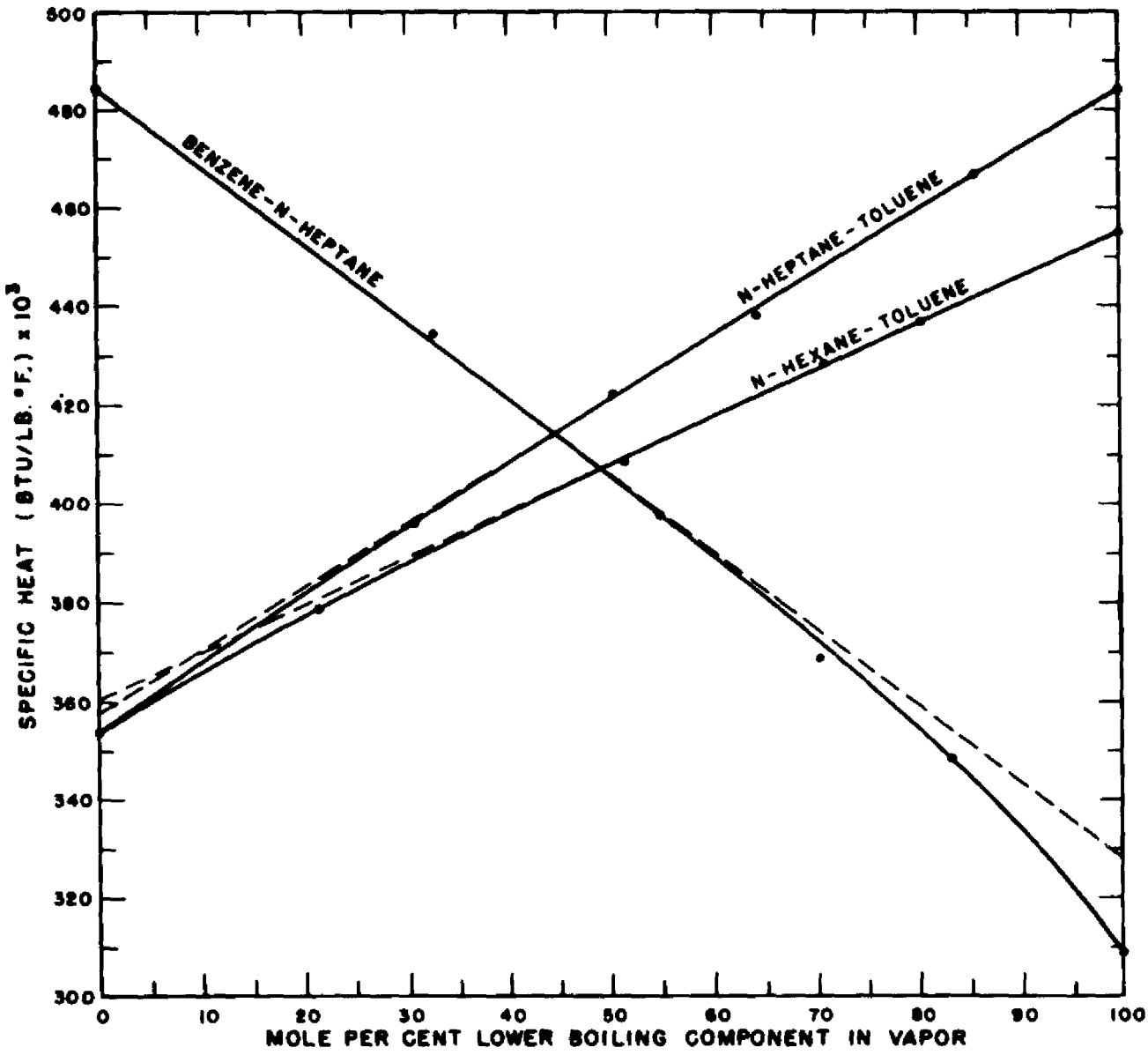

FIGURE 43. SPECIFIC HEAT OF BINARY VAPOR MIXTURES

The resulting equation was tested by the authors on 42 liquids. The average and maximum deviations of the calculated from observed data were 2.6 and -7.4 per cents. The second method proposed by the authors is based on the theorem of corresponding states. The equation developed from this method was also tested by the authors both with respect to thermal conductivity and its temperature coefficients for 47 liquids. The average and maximum deviations of the correlation from their experimental data were 1.5 and +6.6 per cents, respectively.

Reid and Sherwood (79) compared accumulated experimental data with Sakiadis and Coates correlations as well as relationships proposed by other investigators. Their comparison indicated that the method of Sakiadis and Coates, based on the theorem of corresponding states showed the lowest average per cent deviation of 12 per cent, as compared to 13 to 20 per cents determined by other methods. Reid and Sherwood also recommend the Sakiadis and Coates methods for estimating the thermal conductivity of liquids.

Based on Sakiadis and Coates recommendation, and long before Reid and Sherwood recommendation was published, the author calculated, based on method employing the theorem of corresponding states, the thermal conductivity of the liquid hydrocarbons used in this investigation. These data are summarized in Table XIX.

VI-B Pure Vapors. Literature offers few data on thermal conductivity of pure hydrocarbon vapors. McAdams (62) reports the data of Moser (67) for benzene and only at two temperatures for n-heptane. Lamber, et al., (54) report data for benzene at two temperatures. Moser data for benzene

is also reported by Bromley (11,12). Vines and Bennett (109) determined thermal conductivity for benzene and n-hexane. It is reported by Vines and Bennett that their experimental data for the two hydrocarbons and 29 other polar and nonpolar vapors are accurate to ± 1 per cent. The data by various authors are summarized in Table XIX. No data could be found for toluene vapor. For benzene vapor Moser data are about 8 per cent and Lambert, et al., data are less than 2 per cent greater than those reported by Vines and Bennett.

Generalized charts, based on reduced temperature and pressure, for thermal conductivity of gases are prepared by Gamson (36) and Grant (38). These generalized charts may be useful for high pressure-high temperature conditions. However; the reduced pressure parameters used in these charts were not in a low enough range to be useful in this work. Moreover, it is indicated the reliability of the charts is ± 15 per cent. For this type of accuracy, it is much simpler to use a simplified equation (as will be shown later) to determine the thermal conductivity directly, without any need for reduced temperature or pressure calculation.

In the absence of any experimental data for toluene vapor and lack of sufficient data on n-heptane vapor, a study was undertaken to find a method for calculating the thermal conductivity of organic vapors. Of the many correlations presented in literature (11,12,18,31,41-44,61) for calculating the thermal conductivity of gases and vapors at low pressure, reportedly (33,79) the complex equations presented by Bromley (11,12) would yield slightly better accuracy than other correlations. However, time required for evaluating or estimating many physical

TABLE XIX

THERMAL CONDUCTIVITIES OF LIQUID AND VAPOR HYDROCARBONS

Units (Btu/Hr. Ft.^{2} °F./Ft.) x 10^3

Liquids

Temperature °F.	Benzene	n-Heptane	n-Hexane	Toluene
40	97.4	80.5	78.0	100.0
60	94.0	78.0	75.0	97.3
80	91.0	75.3	72.0	94.0
100	88.0	73.0	69.2	91.2
120	84.0	70.3	66.5	88.0
140	81.5	67.5	64.0	85.5
160	78.5	65.0	61.0	82.5
180	75.0	62.5		79.5
200		60.0		76.5
220				74.5

Vapors

Benzene								n-Heptane				n-Hexane				Toluene	
Calculated		Moser		Lambert et al.		Vines and Bennett		Calculated		Moser		Calculated		Vines and Bennett		Calculated	
t, °F.	k	t, °F.	k	t, °F.	k	t, °F.	k	t, °F.	k	t, °F.	k	t, °F.	k	t, °F.	k	t, °F.	k
160	7.07	32	5.05	151	7.79	176	8.46	160	8.01	212	10.3	160	8.89	176	10.43	180	7.30
170	7.29	114	7.11	185	8.87	212	9.57	170	8.68	392	11.2	170	9.13	212	11.65	190	7.53
180	7.51	212	10.30			248	10.70	180	8.89			180	9.37	248	12.93	200	7.76
190	7.74	363	14.80			284	11.89	190	9.10			190	9.62	284	14.31	210	7.99
200	7.96	415	17.12			320	13.10	200	9.32			200	9.86	320	15.74	220	8.22
210	8.20							210	9.55			210	10.12			230	8.50
220	8.43							220	9.77			220	10.38			240	8.70

properties of the compounds and/or the various factors involved in his equations do not justify its use for many engineering applications. Therefore, a much simpler though less accurate equation that was developed by Maxwell (61) and corrected by Encken (31) was used for estimating the thermal conductivity of the hydrocarbons used in this investigation. This equation is presented (33,79) as

$$k = \mu \left(C_p + \frac{2.48}{M}\right) \qquad \text{B-14}$$

where all properties are in engineering units. It is reported (33) that Equation B-14 is applicable with a fair degree of accuracy to any gas at moderate or high temperature, and especially in the 100-1000 mm. mercury range.

The thermal conductivity of benzene, n-heptane, n-hexane, and toluene were calculated with using Equation B-14. These are summarized in Table XIX. The calculated data for benzene, n-hexane, and n-heptane were 14.3, 13.8, and 13.3 per cents lower, respectively, than those reported by Vines and Bennett (109) for benzene and n-hexane, and Moser (67) for n-heptane. Vines and Bennett data for benzene and n-hexane, Moser data for n-heptane, and the calculated data for toluene were used in subsequent calculations.

VI-C Vapor Mixtures. According to Gambill (33) and Reid and Sherwood (79) the most accurate method of calculating the thermal conductivity of vapor mixtures available today is that of Lindsay and Bromley (58). Their correlation for binary vapor mixtures is:

$$k_m = \frac{k_1}{1 + \left(A_{12} \frac{y_2}{y_1}\right)} + \frac{k_2}{1 + \left(A_{21} \frac{y_1}{y_2}\right)} \qquad \text{B-15}$$

where y_1, y_2 are mole fractions of components; and A_{12} is given as

$$A_{12} = 0.25\left(1 + a^{0.5}\right)^2 b \qquad \text{B-16}$$

$$a = \frac{\mu_1}{\mu_2}\left(\frac{M_1}{M_2}\right)^{0.75}\left(\frac{1 + S_1/T}{1 + S_2/T}\right) \qquad \text{B-17}$$

$$b = \frac{1 + S_{12}/T}{1 + S_1/T} \qquad \text{B-18}$$

Here μ, M and S are respectively, viscosity, molecular weight, and Sutherland's constant. T is the absolute temperature in °K and S_{12} is the geometric mean of S_1 and S_2. Subscripts 1 and 2 refer to components 1 and 2. To obtain A_{21} the subscripts are interchanged.

The calculation of Sutherland constants from theory in a rigorously accurate manner has not yet been accomplished; Enskog (30) has given a relation between S and the law force of intermolecular attraction. The most generally useful relation is one originally due to Sutherland (99), later put forward more clearly by Vogel (110); Vogel's equation is

$$S = 1.47\ T_b \qquad \text{B-19}$$

Since S is a measure of the intermolecular potential energy, which in turn determine the ease of vaporization, such a relation appears quite logical. The equation is believed to have a maximum error of ± 10 per cent.

Lindsay and Bromley (58) compared their equation with experimental thermal conductivity for 85 binary mixtures. The average and maximum

per cent deviations from experimental data for these mixtures were 1.9 and +7.3 per cents, respectively.

The thermal conductivity of the binary mixtures for benzene-n-heptane, n-heptane-toluene, n-hexane-toluene systems were calculated, using B-15 through B-19. The calculated data are summarized in Table XX and are shown in Figure 44 as a function of vapor composition incoming to the primary condenser. In the same manner as for the viscosity and specific heat of the binary systems, the lines have slight curvature and again, in each case the curvature of the line starts when the vapor composition is predominantly aromatic. The maximum per cent deviations from straight lines are +3.6, -1.3, and +8.12 per cents for benzene-n-heptane, n-heptane-toluene, and n-hexane-toluene, respectively.

VII PRANDTL NUMBER OF VAPOR MIXTURES

Knowing the specific heat, viscosity, and the thermal conductivity of each vapor mixture, the Prandtl number and $(Pr)^{0.56}$ for the vapor mixtures were calculated. These data are summarized in Table XX.

VIII NUSSELT'S PHYSICAL PROPERTY GROUP

This factor is needed in order to calculate the theoretical film coefficient of condensing pure vapors. Therefore, at a series of assumed film temperature the factor Φ was calculated for each hydrocarbon. These factors are presented in Table XXI and Figure 45.

IX MASS AND HEAT TRANSFER FACTOR

For flow of vapors transverse to cylinder, Treybal (106) has presented a plot of mass and heat transfer factor as a function of vapor Reynolds

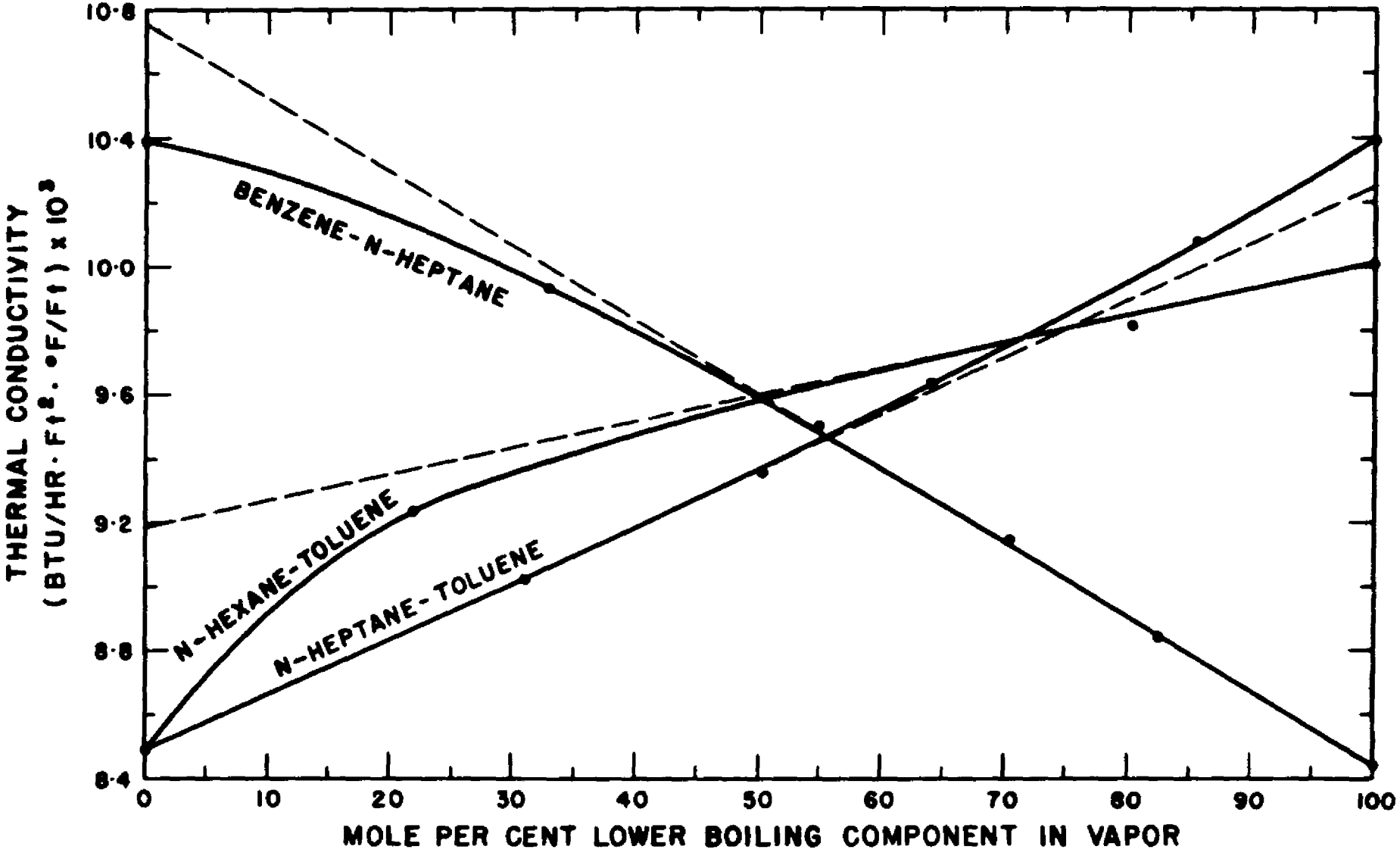

FIGURE 44. THERMAL CONDUCTIVITY OF BINARY VAPOR MIXTURES

TABLE XX

PHYSICAL PROPERTIES OF BINARY VAPOR MIXTURES

Vapor Temperature °F. °F.	Mole Per Cent Lower Boiling Component in Vapor	Density ρ Lbs. / Ft.^{3}	Viscosity μ Lbs. / Hr. Ft.	Diffusivity D_v Ft.^{2} / Hr.	Schmidt Number $\frac{\mu}{\rho D_v}$	$\left(\frac{\mu}{\rho D_v}\right)^{0.56}$	Specific Heat C_p Btu / Lb. °F.	Thermal Conductivity k Btu / Hr. Ft. °F.	Prandtl Number $\frac{C_p \mu}{k}$	$\left(\frac{C_p \mu}{k}\right)^{0.56}$
				I BENZENE - n-HEPTANE SYSTEM						
176.2	100.00	0.1737	0.0217				0.3090	0.0084	0.7983	0.8815
179.8	82.57	0.1824	0.0210	0.1530	0.7527	0.8529	0.3480	0.0088	0.8267	0.8989
184.4	70.37	0.1869	0.0206	0.1553	0.7102	0.8256	0.3687	0.0092	0.8360	0.9047
190.8	54.94	0.1939	0.0202	0.1586	0.6561	0.7898	0.3976	0.0095	0.8486	0.9093
198.7	32.98	0.2033	0.0196	0.1621	0.5935	0.7467	0.4346	0.0099	0.8569	0.9172
209.2	0.00	0.2175	0.0187				0.4845	0.0104	0.8734	0.9270
				II n-HEPTANE - TOLUENE SYSTEM						
209.2	100.00	0.2175	0.0187				0.4845	0.0104	0.8734	0.9270
210.9	85.68	0.2139	0.0192	0.1487	0.6022	0.7528	0.4662	0.0101	0.8857	0.9373
213.8	64.13	0.2078	0.0197	0.1498	0.6348	0.7753	0.4386	0.0096	0.9000	0.9460
217.0	50.32	0.2043	0.0202	0.1513	0.6545	0.7887	0.4212	0.0093	0.9103	0.9508
222.5	31.19	0.1969	0.0210	0.1538	0.6855	0.8094	0.3964	0.0090	0.9215	0.9553
231.2	0.00	0.1910	0.0222				0.3580	0.0085	0.9273	0.9586
				III n-HEXANE - TOLUENE SYSTEM						
155.7	100.00	0.2000	0.0182				0.4555	0.0100	0.8267	0.8990
175.9	80.40	0.1982	0.0190	0.1479	0.6465	0.7833	0.4370	0.0098	0.8482	0.9110
185.1	70.93	0.1975	0.0195	0.1521	0.6688	0.7968	0.4281	0.0097	0.8580	0.9154
201.2	51.30	0.1954	0.0203	0.1595	0.6521	0.7871	0.4088	0.0096	0.8671	0.9232
221.2	21.92	0.1930	0.0215	0.1687	0.6606	0.7928	0.3784	0.0092	0.8809	0.9315
231.2	0.00	0.1910	0.0222				0.3580	0.0085	0.9273	0.9586

TABLE XXI

NUSSELT PHYSICAL PROPERTY GROUP $\Phi = \left(\frac{k_f^3 \rho_f^2 g}{\mu_f^2} \right)^{1/3}$

OF THE HYDROCARBONS

Temperature °F.	Benzene	Toluene	n-Heptane	n-Hexane
120	857.0		748.0	822.5
130	869.7		754.2	825.5
140	883.0	870.0	759.4	827.8
150	895.0	882.5	763.5	829.8
160	907.9	892.5	768.5	
170	920.5	900.6	773.0	
180		908.2	777.0	
190		915.8	781.0	
200		923.5	781.5	
210		931.2		
220		938.7		

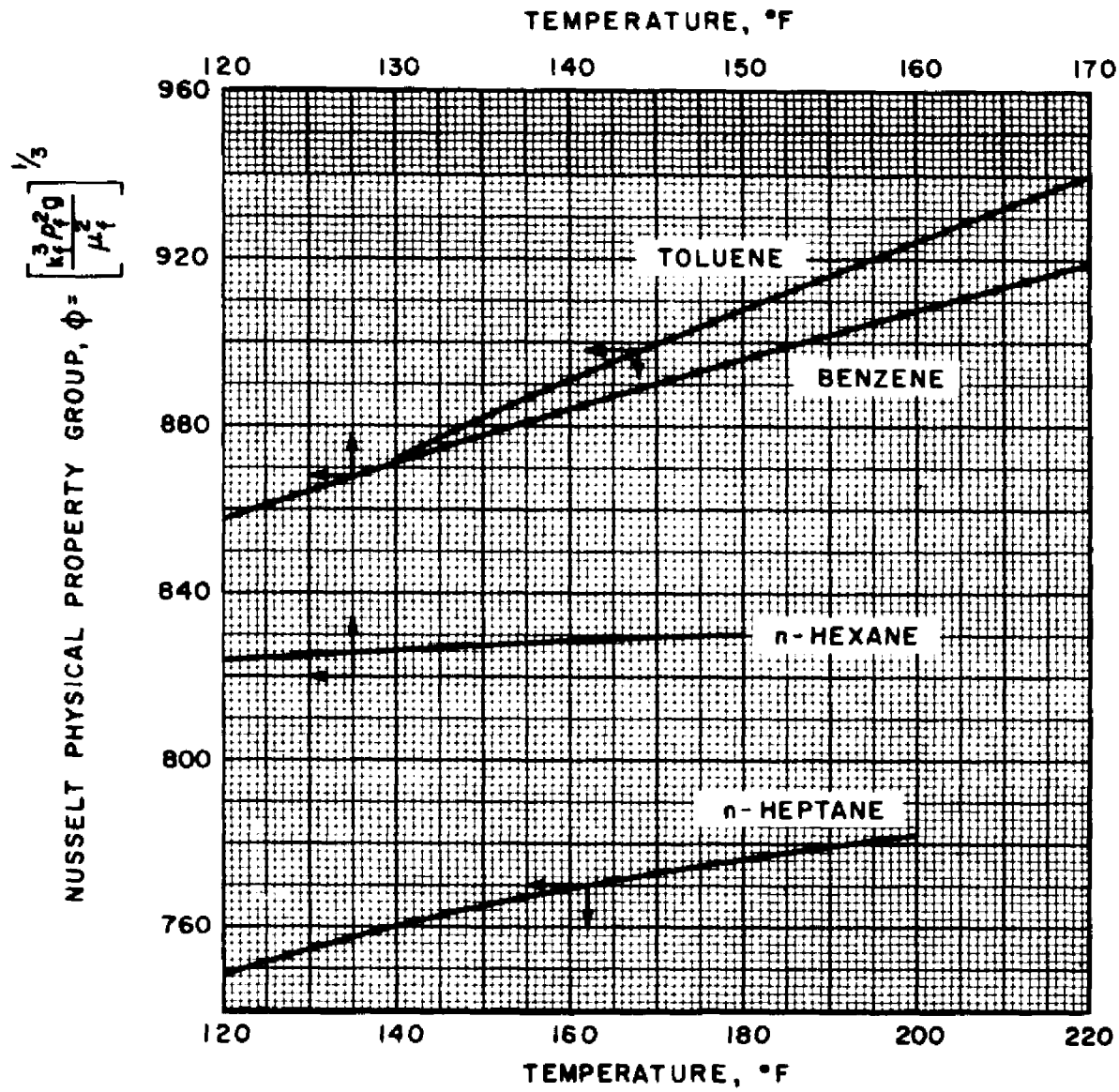

FIGURE 45. NUSSELT PHYSICAL PROPERTY GROUP OF THE HYDROCARBONS

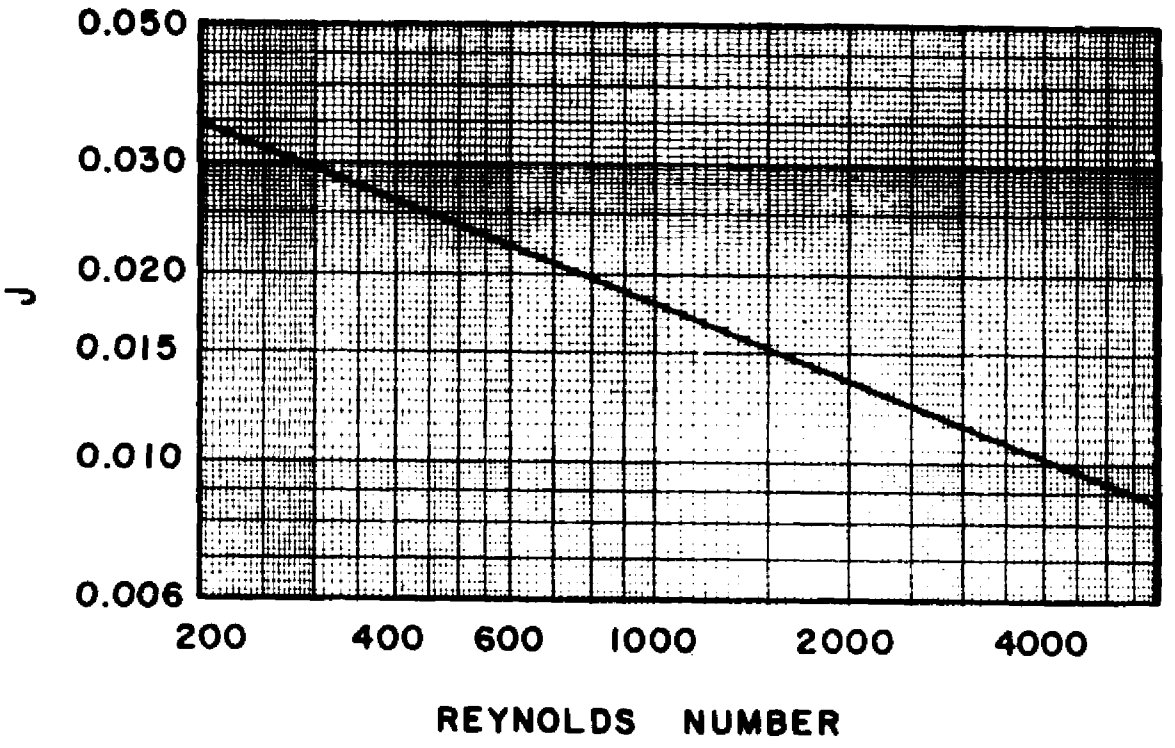

FIGURE 46. MASS AND HEAT TRANSFER CORRELATION FOR FLOW OF GASES TRANSVERSE TO CYLINDERS

number. This is line No. 3, Figure 3.11 of Mass Transfer Operations (106). The line which is reproduced on Figure 46, is a straight line on log-log paper and can be represented by the following equation

$$J = 0.3116 \text{ (Re. No.)}^{-0.413} \qquad \text{B-20}$$

X VAPOR-LIQUID EQUILIBRIUM DATA

Many references (7,14,40,68,83,45,47) are available in literature for vapor-liquid equilibrium data for binary mixtures of paraffins and aromatic hydrocarbons. With minor exceptions, the experimental data of various investigators were found to be very close to each other. For this work, the more recent data of Myers (68,69) for benzene-n-heptane and n-hexane-toluene and the Rose, et al. (83) data for n-heptane-toluene systems were used. These data are shown in Table XXII.

XI ENTHALPY-CONCENTRATION DIAGRAMS

These diagrams were needed to determine the latent of condensation of the binary systems. The diagrams were constructed according to Dodge (25). Since specific heats for these mixtures were not available, an average value was calculated using mole per cent as basis. Latent heat of vaporization of pure hydrocarbons was taken from A.P.I. Project 44. Heats of mixing of benzene-n-heptane are given by Brown, et al. (13) at 20°C. and by Brown and Ewald (14) at 70°C. Mathieson and Thynne (60) report data for n-heptane-toluene and n-hexane-toluene systems 20°C. Tsao and Smith (107) report data for n-heptane-toluene system at 20°C. Schnaible, et al. (92) report data for n-heptane-benzene and n-heptane-toluene systems at 25°C. Each of the above reports contain

TABLE XXII

VAPOR-LIQUID EQUILIBRIUM DATA FOR BINARY MIXTURES AT ONE ATMOSPHERE

Benzene - n-Heptane System			n-Heptane-Toluene System			n-Hexane-Toluene System		
Mole Per Cent Benzene		Temperature	Mole Per Cent n-Heptane		Temperature	Mole Per Cent n-Hexane		Temperature
Liquid	Vapor	°F.	Liquid	Vapor	°F.	Liquid	Vapor	°F.
0	0	209.2	0	0	231.1	0	0	231.2
6.0	11.9	205.3	10	16.60	225.9	4.0	15.0	222.9
11.9	22.1	201.8	20	29.40	222.1	16.9	44.7	204.2
19.8	33.7	197.4	30	40.05	219.0	27.2	58.9	193.9
26.7	41.7	194.1	40	49.70	216.7	36.9	68.2	186.0
35.0	50.6	190.5	50	58.25	214.7	43.7	73.3	181.0
46.7	61.4	186.1	60	66.40	213.1	50.1	77.2	177.2
58.4	70.0	182.6	70	74.40	211.7	62.8	84.0	170.9
63.9	74.1	181.7	80	82.75	210.6	71.3	88.3	167.0
76.3	82.3	179.1	90	91.20	209.8	79.4	91.7	162.9
89.5	91.5	177.2	100	100.00	209.2	88.2	95.5	159.5
95.7	96.2	176.6				96.2	98.6	156.6
98.6	98.7	176.4				100.0	100.0	155.4
100.0	100.0	176.2						

heats of mixing data for many other binary and ternary systems. Although the above data for each binary system is given at a different temperature, they were found to closely check each other on a comparative temperature basis. All the data except those of Tsao and Smith (107) were published recently and the selection of the data were based upon its availability at the time of constructing the enthalpy-concentration diagram for each system. The selected data are reported in Table XXIII.

The calculated enthalpy-concentration data for the three binary systems are shown in Table XXIV.

XII CALIBRATION CURVE FOR ANALYSIS OF SAMPLES

A calibration curve was constructed for each binary system so that the composition of samples taken during the experiment could be determined. To construct the calibration curves, the refractive indices of the pure hydrocarbons and ten known mixtures for each binary system were measured at 25°C. with a Bauch and Lomb Precision Refractometer, Model 33-45-01. The data are reported in Table XXV.

TABLE XXIII

HEATS OF MIXING FOR BINARY SYSTEMS AT ONE ATMOSPHERE

Benzene-n-Heptane at 70°C.
n-Heptane-Toluene at 25°C.
n-Hexane-Toluene at 20°C.

(Units: Btu/Lb.-Mole)

Mole Fraction of the More Volatile Component in the System	Benzene-n-Heptane System	n-Heptane-Toluene System	n-Hexane-Toluene System
0.1	90.99	92.24	72.45
0.2	171.63	175.60	126.62
0.3	239.44	195.70	164.78
0.4	291.26	214.17	188.37
0.5	323.21	215.62	198.00
0.6	330.70	200.70	191.79
0.7	308.47	171.38	167.86
0.8	250.50	128.25	126.82
0.9	150.16	71.36	70.11

TABLE XXIV

ENTHALPY-CONCENTRATION DATA FOR BINARY SYSTEMS AT ONE ATMOSPHERE

(Units: Btu/Lb.)

Mole Fraction of the More Volatile Component In the System	Benzene-n-Heptane System		n-Heptane-Toluene System		n-Hexane-Toluene System	
	h_l	h_g	h_l	h_g	h_l	h_g
0.0	99.50	235.5	87.89	244.2	87.06	243.3
0.1	92.86	234.5	87.66	244.5	80.04	245.0
0.2	86.96	233.5	87.87	243.0	75.62	242.0
0.3	82.03	233.0	88.52	241.5	72.59	239.2
0.4	76.74	232.0	89.53	240.5	70.87	236.0
0.5	72.75	231.5	90.72	239.5	69.63	233.0
0.6	69.25	230.5	92.10	238.5	68.82	229.5
0.7	66.26	230.0	93.64	237.5	68.23	226.5
0.8	63.88	230.0	95.48	237.0	67.80	223.5
0.9	62.45	230.0	97.47	236.5	67.86	220.0
1.0	61.78	231.1	99.35	235.5	68.72	212.7

TABLE XXV

REFRACTIVE INDEX CALIBRATION DATA FOR THE BINARY SYSTEMS

TEMPERATURE 25°C.

Benzene-n-Heptane System		n-Heptane-Toluene System		n-Hexane-Toluene System	
Mole Per Cent Benzene	N_D^{25}	Mole Per Cent n-Heptane	N_D^{25}	Mole Per Cent n-Hexane	N_D^{25}
0	1.38558	0	1.49214	0	1.49214
7.22	1.39012	7.69	1.47870	10.30	1.47691
16.58	1.39666	15.24	1.46894	17.16	1.46748
25.32	1.40321	22.78	1.45919	25.56	1.45624
32.67	1.40898	32.61	1.44650	35.17	1.44375
45.34	1.42073	41.04	1.43760	44.39	1.43229
57.09	1.43229	51.53	1.42625	54.38	1.42041
67.48	1.44464	63.87	1.41461	65.62	1.40799
77.22	1.45797	73.34	1.40603	76.34	1.39607
85.28	1.47039	84.31	1.39689	87.21	1.38482
92.30	1.48194	91.24	1.39125	91.81	1.38017
100.00	1.49744	100.00	1.38558	100.00	1.37225

APPENDIX C

SAMPLE CALCULATIONS

I - HEAT TRANSFER CALCULATIONS FOR PURE HYDROCARBONS

The following sample calculations are made for Test No. 2 of pure benzene series.

I - A. Experimental Condensing Film Coefficient of Heat Transfer. The experimental heat transfer coefficient was calculated by the Newton's equation

$$h = q/A\ \Delta T \qquad 1\text{-C}$$

$$q = W'\lambda \qquad 2\text{-C}$$

$$\Delta T = T_V - T_W \text{ (Average)} \qquad 3\text{-C}$$

Latent heat of condensation for benzene = 169.34 Btu/Lb.

$$W' = 948.4 \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Mins.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} = 125.45 \text{ Lbs/Hr.}$$

$$q = 125.45 \frac{\text{Lbs.}}{\text{Hr.}} \times 169.34 \frac{\text{Btu.}}{\text{Lb.}} = 21{,}244 \text{ Btu/Hr.}$$

$$A = \pi DL = 3.14 \times \frac{3.908}{12} \times \frac{28}{12} = 2.386 \text{ Ft.}^2$$

$$\Delta T = 177.3 - 140.5 = 36.8°\text{F.}$$

Substituting the above figures in Equation 1-C and solving for h,

$$h = 21{,}244 \frac{\text{Btu}}{\text{Hr.}} / (2.386 \text{ Ft.}^2)(36.8°\text{F.}) = 242 \frac{\text{Btu}}{\text{Hr. Ft.}^2 \text{ °F.}}$$

I - B. Theoretical Condensing Film Coefficient of Heat Transfer. The theoretical coefficient is calculated in the following manner

$$h = 1.51\ \phi/(4\Gamma/\mu_f)^{1/3} \qquad 4\text{-C}$$

where $$\phi = (k_f^{\,3}\rho_f^{\,2}g/\mu_f^{\,2})^{1/3} \qquad 5\text{-C}$$

All the physical properties are determined at the film temperature T_F, where

$$T_F = T_V - 0.75\ (T_V - T_W) \qquad 6\text{-C}$$

$$T_F = 177.3 - 0.75\ (177.3 - 140.5) = 149.7°F.$$

From Figure 45, ϕ was read at 149.7°F. to be 894.7.

The Reynolds number of condensate is

$$\frac{4\Gamma}{\mu_f} = \frac{4\left[\frac{948.4\ \text{Gms.}}{\text{Min.}} \times \frac{60\ \text{Min.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6\ \text{Gms.}}\right]}{\frac{28}{12}\ \text{Ft.} \times 2.42\ (0.368\ \text{centiposies})} = 241$$

$$h = 1.51\ (894.7)\ /\ (241)^{1/3} = 217\ \text{Btu/Hr. Ft.}^2\ °F.$$

h/ϕ, Experimental = 242/894.7 = 0.270

h/ϕ, Theoretical = 217/894.7 = 0.243

I - C. Rohsenow's Correction Factor to Condensing Film-Coefficient of Heat Transfer for Pure Hydrocarbons. First the value of

$$C_{pf}\ \Delta T/\lambda, \qquad 7\text{-C}$$

is calculated to see if it is less 0.2

$T_F = 149.7°F.$ $\qquad C_{pf} = 0.4470$ Btu/lb.

$\Delta T = 36.8°F.$ $\qquad \lambda = 169.34$ Btu/lb.

$$C_{pf}\ \Delta T/\lambda = 0.4470\ (36.8)/169.34 = 0.097$$

Since this is smaller than 0.2, the following correction is applied

$$\lambda' = \lambda + 3/8\ (\Delta T\ C_{pf}) \qquad 8\text{-C}$$

$$\lambda' = 169.34 + 3/8\ (36.8 \times 0.4470) = 175.5\ \frac{\text{Btu}}{\text{Lb.}}$$

$$h'_m = h\ (\lambda'/\lambda)^{0.25} \qquad 9\text{-C}$$

$$h'_m = 217\ (175.5/169.34)^{0.25} = 218.9\ \frac{\text{Btu}}{\text{Hr. Ft.}^2\ °\text{F.}}$$

$$\Delta h_m = h'_m - h_m \qquad 10\text{-C}$$

$$\Delta h_m = 218.9 - 217 = 1.9\ \frac{\text{Btu}}{\text{Hr. Ft.}^2\ °\text{F.}}$$

II - HEAT TRANSFER CALCULATIONS FOR MIXED HYDROCARBONS

The following sample calculations are made for Test No. 1 of series

A, for benzene-n-heptane system where the composition of the primary condensate is 82.5 mole per cent benzene in n-heptane.

II - A. Material Balance and Composition of Vapor Entering the Primary Condenser. Let W represent the rate of condensation in the primary condenser in Gms./Min. W_2 represents the rate of condensation in the auxiliary condenser in Gms./Min. Then total vapor entering the primary condenser, W_1, is the sum of W and W_2.

Let y_{A1} and Y_{A1} represent the mole per cent and weight per cent, respectively, of benzene in vapor entering the primary condenser, y_{A2} and Y_{A2} represent the mole per cent and weight per cent of benzene in vapor leaving the primary condenser. y_{A2} is also equal to x_{A2}, mole per cent of benzene in the condensate from the auxiliary condenser. Therefore, Y_{A2} is equal to X_{A2}, weight per cent benzene in auxiliary condensate. Let x_A and X_A represent mole per cent and weight per cent, respectively, of benzene in the primary condensate. Let also the ratio of W/W_2 be represented by R.

$$W_1 = W + W_2 \qquad 11\text{-C}$$

$$W/W_2 = R \text{ or } W = W_2 R \qquad 12\text{-C}$$

$$W_1 = W_2 (R + 1) \qquad 13\text{-C}$$

Then, material balance for benzene is:

$$Y_{A1} (W_1) = X_A (W) + Y_{A2} (W_2) \qquad 14\text{-C}$$

Substituting Equations 12-C and 13-C in the above equation and simplifying would give

$$Y_{A1}\ (R + 1) = Y_{A2} + R\ (X_A) \qquad 15\text{-C}$$

Solving for Y_{A1},

$$Y_{A1} = \frac{Y_{A2} + R\ (X_A)}{R + 1} \qquad 16\text{-C}$$

$$W = 276.7\ \frac{\text{Gms.}}{\text{Min.}},\ W_2 = 13.0\ \frac{\text{Gms.}}{\text{Min.}},\ R = \frac{276.7}{13.0} = 21.28$$

From Table IV - A,

y_{A2} = 85.20 mole per cent benzene

Y_{A2} = 81.77 weight per cent benzene

x_A = 82.60 mole per cent benzene

X_A = 78.72 weight per cent benzene

$$Y_{A1} = \frac{81.77 + 21.28\ (78.72)}{22.28} = 78.85 \text{ weight per cent}$$

y_{A1} = 82.72 mole per cent

$y_{A1} - x_A = 82.72 - 82.60 = 0.12$ mole per cent

II - B. Murphree Plate Efficiency. The following equation is used in literature (80,106) to determine the plate efficiency in distillation columns.

$$E^{\circ}_{MV} = \frac{y_{A1} - y_{A2}}{y_{A1} - y_{A}^{*}} \qquad 17\text{-C}$$

However, here it can be used to determine the degree of enrichment of the residual vapors. Substituting the composition of the various vapor streams in Equation 17-C,

$$E^{\circ}_{MV} = \frac{82.72 - 85.20}{82.72 - 86.60} \ (100) = 63.9 \text{ per cent}$$

shows that the enrichment of the residual vapors was 63.9 per cent of that predicted from the temperature-composition diagram for the condensate composition.

II - C. Interfacial Composition and Temperature. Equation 49 which is renumbered here as Equation 18-C, is used to determine the interfacial composition

$$\ln \frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} = \frac{\ln (V_1/V_2)}{1 - (V_1/V_2)^{-\frac{Hg}{h}}} \qquad 18\text{-C}$$

II - C - 1. Vapor Flow Rates. Let V_1 and V_2 represent vapor flow rates entering and leaving the primary condenser.

Molecular weight of incoming vapor = 81.96 $\frac{\text{Lbs.}}{\text{Lb.-Mole}}$

$$W_1 = W + W_2 = 276.7 + 13 = 289.7 \ \frac{\text{Gms.}}{\text{Min.}}$$

$$V_1 = 289.7 \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Mins.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} \times \frac{\text{Lb.-Mole}}{81.96 \text{ Lbs.}}$$

$$= 0.4675 \text{ Lb.-Moles/Hr.}$$

$$\text{Molecular weight of vapor leaving} = 81.40 \frac{\text{Lbs.}}{\text{Lb.-Mole}}$$

$$V_2 = 13.0 \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Mins.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} \times \frac{\text{Lb.-Mole}}{81.40 \text{ Lbs.}}$$

$$= 0.0211 \text{ Lb.-Moles/Hr.}$$

$$V_1/V_2 = 0.4675/0.0211 = 22.16$$

II - C - 2. Mass Velocities of Entering Vapor. Let G_{M1} and G_1' represent mass velocities of incoming vapor to the primary condenser in Lb.-Moles/Hr. Ft.^{2} and Lbs./Hr. Ft.^{2}, respectively.

Assuming that one-half of total vapor entering the jacket flows through each section of the annulus between the jacket and the condensing bar, then cross sectional area of the flow for each section is calculated as follows

$$S = L (R_2 - R_1) \qquad 19\text{-C}$$

where L is the effective length of condenser = 28 inches.

R_2 is inside radius of the jacket = $\frac{6.025}{2}$ inches.

R_1 is outside radius of the condenser = $\frac{3.908}{2}$ inches.

Therefore,

$$S = \frac{28}{12}\left(\frac{6.025 - 3.908}{12 \times 2}\right) = 0.2058 \text{ Ft.}^2$$

The equivalent diameter for each section may be written as

$$D_e = \frac{4L\,(R_2 - R_1)}{2L + 2\,(R_2 - R_1)} \qquad 20\text{-C}$$

However, since the end effects $2(R_2 - R_1)$ are small compared to 2L, these can be neglected and Equation 20-C reduces to

$$D_e = \frac{4L\,(R_2 - R_1)}{2L} = D_2 - D_1 \qquad 21\text{-C}$$

$$D_e = (6.025 - 3.908)/12 = 0.1764 \text{ Ft.}$$

Total vapor flow to the primary condenser is

$$W_1 = W + W_2 = 276.7 + 13 = 289.7 \text{ Gms./Min.}$$

$$G_1' = \frac{289.7}{2}\frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Mins.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} \times \frac{}{0.2058 \text{ Ft.}^2}$$

$$= 93.1 \text{ Lbs./Hr. Ft.}^2$$

$$G_{M1} = 93.1 \text{ Lbs./Hr. Ft.}^2/81.96 \text{ Lbs./Lb.-Mole}$$

$$= 1.1359 \text{ Lb.-Moles/Hr. Ft.}^2$$

II - C - 3. Reynolds Number of Entering Vapor

$$\text{Re. No.} = D_e \, G_1' / \mu_m$$

The viscosity of entering vapor mixture is read from Table XX to be 0.0210 Lbs./Hr. Ft.

$$\text{Re. No.} = (0.1764)(93.1)/(0.0210) = 780$$

II - C - 4. Mass Transfer Coefficient of Entering Vapor. Knowing the Reynolds number of vapor, J_D factor is read from Figure 46 to be 0.0200.

The mass transfer coefficient, k_{g1}, is calculated from

$$k_{g1} = J_D \, (G_{M1})/P \, (Sc)^{0.56} \qquad 22\text{-C}$$

Since the experiment was performed slightly above (2 to 3 inches of water) atmospheric pressure, then P in Equation 22-C can be taken as one atmosphere.

The Schmidt number and $(Sc)^{0.56}$ for the vapor entering the primary condenser are read from Table XX, to be 0.7527 and 0.8529, respectively.

Substituting the above determined figures in Equation 22-C, and solving for k_{g1}

$$k_{g1} = 0.020 \, (1.1359)/(1)(0.8529) = 0.0266 \, \frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Atm.}}$$

Since in subsequent calculations k_{g1} is required in terms of

$\frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Mole Fraction}}$ rather than in terms of $\frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Atm.}}$,

and since the experiment was performed at essentially one atmosphere, then k_{g1}, determined above, can be written in term of mole fraction driving force,

$$k_{g1} = 0.0266 \frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Mole Fraction}}$$

II - C - 5. Height Equivalent to a Transfer Unit, H_{g1}. Equation 46 which is renumbered here as Equation 23-C is

$$H_{g1} = G_{M1}/k_{g1}\, a\, P \qquad 23\text{-C}$$

Here P is one atmosphere and a is the interfacial area per unit volume of apparatus.

$$a = \frac{A}{V}\,, \qquad 24\text{-C}$$

where A is the total surface area of condenser and V is the total volume of the condensing section in the apparatus. For flow of vapor through one-half of the annulus between the jacket and the condenser, Equation 24-C can be written as

$$a = \frac{\pi R_1 L}{\frac{\pi}{2}(R_2^2 - R_1^2) L} = \frac{2R_1}{R_2^2 - R_1^2} \qquad 25\text{-C}$$

$$a = \frac{2\ (3.908/2 \times 12)}{(6.025/2 \times 12)^2 - (3.908/2 \times 12)^2} = 8.920\ \text{Ft.}^2/\text{Ft.}^3$$

Substituting the values of G_{Ml}, k_{gl} and **a** in Equation 23-C and solving for H_{gl},

$$H_{gl} = 1.1359/(0.0266)(8.920) = 4.78\ \text{Ft.}$$

Equations 22-C and 23-C may be combined to give

$$H_{gl} = \frac{(Sc)^{0.56}}{J_D\ a} \qquad 26\text{-C}$$

For flow of vapors transverse to single cylinder, J_D can be expressed as a function of Reynolds number of vapor. This is shown in Appendix B, Section IX as

$$J_D = 0.3116\ (\text{Re.No.})^{-0.413} \qquad 27\text{-C}$$

Substituting this equation for J_D in Equation 26-C, would result in

$$H_{gl} = \frac{1}{0.3116\ a}\ (\text{Re.No.})^{0.413}\ (Sc)^{0.56} \qquad 28\text{-C}$$

For the present equipment **a** is determined (Section II-C-5) to be 8.92 $\text{Ft.}^2/\text{Ft.}^3$. Applying this to Equation 28-C, and simplifying would give

$$H_{gl} = 0.36\ (\text{Re.No.})^{0.413}\ (\text{Sc})^{0.56} \qquad 29\text{-C}$$

Substituting the values of Reynolds number and Schmidt number determined before in above equation and solving for H_{gl},

$$H_{gl} = 0.36\ (780)^{0.413}\ (0.7527)^{0.56} = 4.80\ \text{Ft.}$$

which is very close to that determined from Equation 23-C.

II - C - 6. Height of Vapor Travel in the Jacket. The height that one-half of vapor travels in each section of annulus in the jacket, h, is determined in the following manner (76).

$$H = \frac{V}{S}\ \frac{\frac{\pi}{2}\ (R_2^2 - R_1^2)\ L}{L\ (R_2 - R_1)} = \frac{\pi}{2}\ (R_2 + R_1) \qquad 30\text{-C}$$

$$H = \frac{\pi}{2} \times \left(\frac{6.025}{2 \times 12} + \frac{3.908}{2 \times 12} \right) = 0.65\ \text{Ft.}$$

II - C - 7. Now H_{gl}/h and $(V_1/V_2)^{-H_{gl}/h}$ can be calculated,

$$\frac{H_{gl}}{h} = \frac{4.78}{0.65} = 7.35$$

and $(V_1/V_2)^{-H_{gl}/h} = (22.16)^{-7.35} = 1.29 \times 10^{-10}$.

It is noticed that the magnitude of $(V_1/V_2)^{-H_{gl}/h}$ is small, therefor the denominator of the right-side of Equation 18-C can be taken as unity.

In that case, as it was pointed out in Chapter II, Equation 49 simplifies to Equation 52.

II - C - 8. Interfacial Composition and Temperature. Equation 52 is renumbered here as Equation 31-C.

$$\frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} = \frac{V_1}{V_2} \qquad 31\text{-C}$$

y_{A1} = 82.72 mole per cent benzene

y_{A2} = 85.20 mole per cent benzene

$\frac{V_1}{V_2}$ = 22.16

Substituting these figures in Equation 31-C and solving for y_{AI}, y_{AI} is 85.33 mole per cent benzene. The interfacial temperature and x_{AI}^*, in equilibrium with y_{AI}, are read, from the temperature-composition diagram, to be:

T_I = 178.4°F.

x_{AI}^* = 81.1 mole per cent benzene.

II - D. Experimental Condensing Film-Coefficients of Heat Transfer. The experimental heat transfer coefficients are calculated using Equations 1-C, 2-C, and 3-C. The coefficients are based on vapor temperature, dew point of the condensate, bubble point of the mixture, or the interfacial temperature minus average wall temperature.

II - D - 1. Rate of Condensation in the Primary Condenser. The rate of condensation is,

$$W' = 276.7 \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Mins.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} = 36.60 \text{ Lbs./Hr.}$$

II - D - 2. Total Heat Transferred. The latent heat of condensation for the binary mixtures, which is the difference between the enthalpy of vapor at the dew point, h_g, and the enthalpy of the liquid of the same composition at the bubble point, h_l, was read from a previously prepared enthalpy-concentration diagram. For the mixture of 82.5 mole per cent benzene in n-heptane, the difference between h_g and h_l at one atmosphere, is 166.5 Btu/Lb. Therefore,

$$q = 36.60 \frac{\text{Lbs.}}{\text{Hr.}} \times 166.5 \frac{\text{Btu}}{\text{Lb.}} = 6094 \text{ Btu/Hr.}$$

II - D - 3. Film-Coefficient of Heat Transfer. Temperature difference based on the vapor temperature is

$$\Delta T_V = T_V - T_W \text{ (Average)}$$

$$\Delta T_V = 179.8 - 173.9 = 5.9°\text{F.}$$

$$h_V = 6094 \frac{\text{Btu}}{\text{Hr.}} / (2.386 \text{ Ft.}^2)(5.9°\text{F.}) = 433 \frac{\text{Btu}}{\text{Hr. Ft.}^2 \text{ °F.}}$$

Temperature difference based on dew point of the mixtures is

$$\Delta T_{DP} = T_{DP} - T_W \text{ (Average)}$$

$$\Delta T_{DP} = 179.0 - 173.9 = 5.1°F.$$

$$h_{DP} = 6094 \frac{Btu}{Hr.} / (2.386 Ft.^2)(5.1°F.) = 505 \frac{Btu}{Hr. Ft.^2 °F.}$$

Temperature difference based on boiling point of the mixture is

$$\Delta T_{BP} = T_{BP} - T_W \text{ (Average)}$$

$$\Delta T_{BP} = 178.2 - 173.9 = 4.3°F.$$

$$h_{BP} = 6094 \frac{Btu}{Hr.} /(2.386 Ft.^2)(4.3°F.) = 608 \frac{Btu}{Hr. Ft.^2 °F.}$$

The latent heat of condensation, λ_I, at x_{AI}^* equal to 81.1 mole per cent benzene is read from enthalpy-concentration diagram to be 166 Btu/Lb. Therefore,

$$q = 36.6 \times 166 = 6075 \text{ Btu/Hr.}$$

The temperature difference is,

$$\Delta T_I = 178.4 - 173.9 = 4.5°F.$$

$$h_I = 6075 \frac{Btu}{Hr.} / (2.386 Ft.^2)(4.5°F.) = 566 \frac{Btu}{Hr. Ft.^2 °F.}$$

II - E. <u>Rate of Sensible Heat Transfer, Heat of Condensation Evolved at the Interface, and the Total Heat Transferred</u>.

II - E - 1. Rate of Sensible Heat Transfer. The rate of arrival of sensible heat by vapor at the interface is given by Equation 20. This equation is renumbered here as Equation 32-C

$$q_s = h_g (T_V - T_I) C_o / 1 - e^{- C_o} \qquad 32\text{-}C$$

where $$C_o = G C_{pf} / h_g \qquad 33\text{-}C$$

Gas-Film Coefficient of Heat Transfer. The gas-film coefficient of heat transfer is determined from the following equation

$$h_g = J_H C_{pf} (G) / (Pr)^{0.56} \qquad 34\text{-}C$$

To determine J_H, heat transfer factor, need to know the Reynolds number of condensing vapor in the primary condenser

$$\text{Re. No.} = De\ G' / \mu_m \qquad 35\text{-}C$$

The equivalent diameter is 0.1764 Ft. and one-half of the cross sectional area of vapor flow is 0.2058 Ft.^{2}. Method of calculation for these terms are shown in Section II-C-2.

$$G' = \frac{276.7}{2} \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Mins.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} \times \frac{}{0.2058 \text{ Ft.}^2}$$

$$= 88.922 \text{ Lbs./Hr. Ft.}^2$$

The viscosity of condensing vapor in the jacket can be taken as the viscosity of the incoming vapor to the primary condenser, without

appreciable error. From Table XX, viscosity of the incoming vapor is 0.0210 Lbs./Hr. Ft.

$$\text{Re. No.} = (0.1764)(88.922)/(0.021) = 750$$

At a Reynolds number of 750, J_H is read from Figure 46 to be 0.0205.

In Equation 34-C, C_{Pf} and $(Pr)^{0.56}$ are to be determined at the vapor-film temperature. The temperature of the main vapor steam is 179.8°F. and the interfacial temperature was determined to be 178.4°F. The temperature of the vapor-film is between these two temperatures. However, even if we assume that the temperature of vapor-film is 178.4°F., the temperature difference of 179.8 - 178.4 = 1.4°F. is not going to affect the specific heat, or the other physical properties of the vapor mixture, greatly. Therefore, for all practical purposes, we can assume that the physical properties of vapor mixture entering the primary condenser can be used here, without any appreciable error. The specific heat and $(Pr)^{0.56}$ for the vapor mixture are read from Table XX to be 0.3480 Btu/Lb. °F. and 0.8989, respectively.

The mass velocity of condensing vapor in the primary condenser is,

$$G = 276.7 \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Min.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} \times \frac{}{2.386 \text{ Ft.}^2}$$

$$= 15.340 \text{ Lbs./Hr. Ft.}^2$$

Substituting the above figures in Equation 34-C and solving for h_g,

$$h_g = \frac{0.0205\ (0.3480)\ 15.340}{0.8989} = 0.1217 \text{ Btu/Hr. Ft.}^2 \text{ °F.}$$

II - E - 2. Rate of Sensible Heat Transfer. Equation 32-C is solved for C_o,

$$C_o = 15.340\ (0.3480)/0.1217 = 43.86$$

Knowing h_g, the vapor and interface temperatures and C_o, Equation 32-C is solved for q_s,

$$q_s = 0.1217\ (179.8 - 178.4)\ 43.86/1 - e^{-43.86}$$

$$= 8.0 \text{ Btu/Hr. Ft.}^2$$

II - E - 3. Latent Heat of Condensation Evolved at the Interface. x^*_{AI} was determined to be 81.1 mole per cent benzene. From enthalpy-concentration diagram, the latent heat of condensation was ready to be 166 Btu/Lb.

$$q_{\lambda I} = G\lambda_I \qquad 36\text{-C}$$

$$q_{\lambda I} = (15.34)(166) = 2546 \text{ Btu/Hr. Ft.}^2$$

On Btu/Hr. basis this is equivalent to 6075 Btu/Hr., which was calculated in Section II-D-3.

II - E - 4. Total Heat Transfer. The total heat transferred is:

$$Q = q_s + q_{\lambda I} \qquad 37\text{-C}$$

$$Q = 8 + 2546 = 2554 \text{ Btu/Hr. Ft.}^2$$

III - MASS TRANSFER CALCULATIONS FOR MIXED HYDROCARBONS

The following sample calculations are made for Test No. 1 of series A, for benzene-n-heptane system where the composition of the primary condensate is 82.5 mole per cent benzene in n-heptane.

III - A. Gas Phase Mass Transfer Coefficient, Number of Individual Gas-Phase Transfer Units, and the Height of an Individual Gas-Phase Transfer Unit.

III - A - 1. Gas-Phase Mass Transfer Coefficient. Equation 53 renumbered here as Equation C-38, can be solved for gas-phase mass transfer coefficient.

$$V_2 (y_{A1} - y_{A2}) = k_g A (y_A - y_{AI})_{1.m.} \qquad 38\text{-C}$$

$$(y_A - y_{AI})_{1.m.} = \frac{(y_{A1} - y_{AI}) - (y_{A2} - y_{AI})}{\ln \dfrac{y_{A1} - y_{AI}}{y_{A2} - y_{AI}}} \qquad 39\text{-C}$$

Substituting the values of y_{A1}, y_{A2}, and y_{AI} from Sections II-A and II-C-8, and solving Equation 39-C for $(y_A - y_{AI})_{1.m.}$,

$$(y_A - y_{AI})_{1.m.} = \frac{(82.72 - 85.33) - (85.20 - 85.33)}{\ln \dfrac{82.72 - 85.33}{85.20 - 85.33}}$$

$$= - 0.827$$

V_2 is 0.0211 Lb.-Moles/Hr.

A is 2.386 Ft.^{2}

Substituting these figures in Equation 38-C,

$$\frac{0.0211}{2} (82.72 - 85.20) = k_g \left(\frac{2.386}{2}\right)(-0.827)$$

and solving for k_g,

$$k_g = 0.0265 \frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Mole Fraction}}$$

k_{gl} obtained, independently, in Section II-C-4, is equal to

$$0.0266 \frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Mole Fraction}}$$

II - A - 2. Number of Individual Gas-Phase Transfer Units. The number of individual gas-phase transfer units can be determined from the following equation

$$N_g = \int_{y_{A1}}^{y_{A2}} \frac{d\, y_A}{y_{AI} - y_A} = \ln \frac{y_{AI} - y_{A1}}{y_{AI} - y_{A2}} \qquad 40\text{-C}$$

$$N_g = \ln \frac{85.33 - 82.72}{85.33 - 85.20} = 2.996$$

III - A - 3. Height of an Individual Gas-Phase Transfer Unit. Equation 55, here called as Equation 41-C, is used to determine the height of an individual gas-phase transfer unit

$$h = H_g N_g \quad \text{or} \quad H_g = h/N_g \qquad 41\text{-C}$$

In Section II-C-6, h was found to be = 0.65 Ft. Therefore,

$$H_g = 0.65/2.996 = 0.217 \text{ Ft.}$$

This also can be determined from Equation 56,

$$H_g = \frac{G_{M2}}{k_g a P} \qquad 42\text{-C}$$

$$\text{where } G_{M2} = \frac{V_2}{2}\,\frac{1}{S} \qquad 43\text{-C}$$

$V_2 = 0.0211$ Lb.-Moles/Hr. (Section II-B-1)

$S = 0.2058$ Ft.^{2} (Section II-C-2)

$$G_{M2} = \frac{0.0211}{2(0.2058)} = 0.0512 \frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2}$$

$a = 8.920$ Ft.^{2}/Ft.^{3} (Section II-C-5)

P = One atmosphere

Substituting these figures in Equation 42-C,

$$H_g = \frac{0.0512}{(0.0265)(8.920)} = 0.216 \text{ Ft.}$$

It is indicated in Section II-C-5 that Equation 29-C is for flow of

gases transverse to cylinders and is not applicable to vapors which are already past the cylinder. However, as a matter of curiosity and for comparison, H_g calculated above will be determined again by using Equation 29-C,

$$H_g = 0.36\ (\text{Re.No.})^{0.413}\ (\text{Sc})^{0.56} \qquad 29\text{-C}$$

$$\text{Re. No.} = \frac{D_e\ G'_2}{\mu_m} \qquad 44\text{-C}$$

Rate of vapor flow from the primary condenser is 13 Gms./Min. Viscosity of vapor leaving the primary condenser is read from Figure 38 to be 0.0211 Lbs./Hr. Ft. Equivalent diameter, D_e, and the cross sectional area of vapor flow are 0.1764 Ft. and 0.2058 Ft.^{2}, respectively (Section II-C-2). Substituting these figures in Equation 44-C, Reynolds number is calculated to be 34.9.

The composition of outgoing vapor from the primary condenser is slightly different from that entering to it. Nevertheless, the $(Sc)^{0.56}$ already calculated based on composition of incoming vapor can be used for the vapor leaving the condenser, without any serious error. From Table XX, $(Sc)^{0.56}$ is read to be 0.8529

$$H_g = 0.36\ (34.9)^{0.413}\ (0.8529) = 1.33\ \text{Ft.}$$

This is about six times greater than those determined from either Equation 41-C or Equation 42-C.

III - B. Over-All Mass Transfer Coefficient, Number of Over-All Gas-Phase Transfer Units, and the Height of the Over-All Gas-Phase Transfer Unit.

III - B - 1. Over-All Mass Transfer Coefficient. Equation 58 can be solved for the over-all mass transfer coefficient.

$$V_2 (y_{A1} - y_{A2}) = K_g A (y_A - y_A^*)_{l.m.} \qquad 45\text{-C}$$

$$(y_A - y_A^*)_{l.m.} = \frac{(y_{A1} - y_A^*) - (y_{A2} - y_A^*)}{\ln \frac{y_{A1} - y_A^*}{y_{A2} - y_A^*}} \qquad 46\text{-C}$$

Substituting the values of y_{A1}, y_{A2}, and y_A^*, from Section II-A, and Table VII-A, in Equation 46-C, and solving for $(y_A - y_A^*)_{l.m.}$,

$$(y_A - y_A^*)_{l.m.} = \frac{(82.72 - 86.60) - (85.20 - 86.60)}{\ln \frac{82.72 - 86.60}{85.20 - 86.60}}$$

$$= - 2.43$$

Substituting the values of V_2, A, and $(y_A - y_A^*)_{l.m.}$ in Equation 45-C and solving for K_g

$$\frac{0.0211}{2} (82.72 - 85.20) = K_g \left(\frac{2.386}{2}\right)(- 2.43)$$

$$K_g = 0.0090 \frac{\text{Lb.-Moles}}{\text{Hr. Ft.}^2 \text{ Mole Fraction}}$$

III- B - 2. Number of Over-All Gas-Phase Transfer Units. The number of over-all gas-phase transfer units can be calculated from the following integral

$$N_{og} = \int_{y_{A1}}^{y_{A2}} \frac{d\,y_A}{y_A^* - y_A} = \ln \frac{y_A^* - y_{A1}}{y_A^* - y_{A2}} \qquad 47\text{-C}$$

$$N_{og} = \ln \frac{86.60 - 82.72}{86.60 - 85.20} = 1.018$$

III - C. Liquid-Phase Mass Transfer Coefficient and the Height of an Individual Liquid-Phase Transfer Unit.

III - C - 1. Liquid-Phase Mass Transfer Coefficient. This coefficient may be determined from the following equation

$$\frac{1}{K_g} = \frac{1}{k_g} + \frac{m}{k_l} \qquad 50\text{-C}$$

$K_g = 0.0090$ Lb.-Moles/Hr. Ft.^{2} Mole Fraction

$k_g = 0.0265$ Lb.-Moles/Hr. Ft.^{2} Mole Fraction

The slope of the equilibrium line at $x_A = 82.6$ mole per cent and $y_A^* = 86.6$ mole per cent is

$$m = \frac{d\,y_A^*}{d\,x_A} \qquad 51\text{-C}$$

$$m = \frac{87.4 - 86.0}{84.0 - 82.1} = 0.737$$

$$\frac{1}{0.0090} = \frac{1}{0.0265} + \frac{0.737}{k_1}$$

$$k_1 = 0.010 \text{ Lb.-Moles/Hr. Ft.}^2 \text{ Mole Fraction}$$

III - C - 2. Height of an Individual Liquid-Phase Transfer Unit. The following equation may be employed for determing the height of an individual liquid-phase transfer unit

$$H_{og} = H_g + m \frac{G_{M2}}{L_M} \qquad 52\text{-C}$$

$H_g = 0.217$ Ft. Section III-A-3

$H_{og} = 0.638$ Ft. Section III-B-3

$m = 0.737$ Section III-C-1

$G_{M2} = 0.0512$ Lb.-Moles/Hr. Ft.^{2} Section III-A-3.

Molecular weight of condensate = 81.96

$$L_M = \frac{276.7}{2} \frac{\text{Gms.}}{\text{Min.}} \times \frac{60 \text{ Min.}}{\text{Hr.}} \times \frac{\text{Lb.}}{453.6 \text{ Gms.}} \times \frac{\text{Lb.-Moles}}{81.96 \text{ Lbs}} \times \frac{}{0.2058 \text{ Ft.}^2}$$

$$= 1.085 \text{ Lb.-Moles/Hr. Ft.}^2$$

$$0.638 = 0.217 + 0.737 \left(\frac{0.0512}{1.085} \right) H_1$$

$$H_1 = 12.1 \text{ Ft.}$$

However, if the surface area of condenser is used in determining L_M, rather than the cross sectional area of vapor flow, H_1 would reduce to 2.1 Ft.

AUTOBIOGRAPHY

Habib Labbauf was born in Hamadan, Iran on September 12, 1924. He entered American Missionary High School, Hamadan, Iran, in September, 1938, and graduated in May, 1944. He attended Abadan College, Abadan, Iran in September, 1944, and graduated in May, 1949, with a Bachelor of Science degree in Petroleum Chemistry. While in college, he also worked in Abadan Refinery, of the former Anglo-Iran Oil Co., Ltd., Abadan, Iran. From June, 1949 until December, 1950 he was in charge of the Lube Oil and Black Oils in the Chemical and Development Department of the above company in Abadan, Iran. In February, 1951 he attended the Graduate School of the University of Tulsa, Tulsa, Oklahoma, and in June, 1953 he received a Master of Science degree in Petroleum Refinery Engineering. In September, 1953 he enrolled in the Graduate School at the Louisiana State University as a Graduate Teaching Assistant in the Chemical Engineering Department. In May, 1955 he received a Master of Science degree in Chemical Engineering. He was a recipient of the Dow Chemical Company fellowship for the year 1955-1956. In June, 1956 he joined the Research and Development Department of the Standard Oil Company (Indiana) where he is active in the Process Design and Economics Division. At the present time he is a candidate for the degree of Doctor of Philosophy in Chemical Engineering.

EXAMINATION AND THESIS REPORT

Candidate: Habib Labbaf

Major Field: Chemical Engineering

Title of Thesis: Heat and Mass Transfer Data for Upflowing Air and Steam Mixtures in Vertical Tubes

Approved:

Major Professor and Chairman

Dean of the Graduate School

EXAMINING COMMITTEE:

Date of Examination:

April 7, 1955